VOLUME 3

PETROLEUM REFINING

DESIGN AND APPLICATIONS HANDBOOK

Mechanical Seperations · Distillation · Packed Towers · Liquid-Liquid Extraction · Process Safety Incidents

A. KAYODE COKER

Scrivener Publishing

WILEY

Petroleum Refining Design and Applications Handbook

Volume 3

Scrivener Publishing
100 Cummings Center, Suite 541J
Beverly, MA 01915-6106

Publishers at Scrivener
Martin Scrivener (martin@scrivenerpublishing.com)
Phillip Carmical (pcarmical@scrivenerpublishing.com)

Petroleum Refining Design and Applications Handbook

Volume 3

- Mechanical Separations
- Liquid-Liquid Extraction
- Distillation
- Packed Towers
- Process Safety Incidents

A. Kayode Coker

Scrivener Publishing

WILEY

This edition first published 2022 by John Wiley & Sons, Inc., 111 River Street, Hoboken, NJ 07030, USA and Scrivener Publishing LLC, 100 Cummings Center, Suite 541J, Beverly, MA 01915, USA
© 2022 Scrivener Publishing LLC
For more information about Scrivener publications please visit www.scrivenerpublishing.com.

All rights reserved. No part of this publication may be reproduced, stored in a retrieval system, or transmitted, in any form or by any means, electronic, mechanical, photocopying, recording, or otherwise, except as permitted by law. Advice on how to obtain permission to reuse material from this title is available at http://www.wiley.com/go/permissions.

Wiley Global Headquarters
111 River Street, Hoboken, NJ 07030, USA

For details of our global editorial offices, customer services, and more information about Wiley products visit us at www.wiley.com.

Limit of Liability/Disclaimer of Warranty
While the publisher and authors have used their best efforts in preparing this work, they make no representations or warranties with respect to the accuracy or completeness of the contents of this work and specifically disclaim all warranties, including without limitation any implied warranties of merchant-ability or fitness for a particular purpose. No warranty may be created or extended by sales representatives, written sales materials, or promotional statements for this work. The fact that an organization, website, or product is referred to in this work as a citation and/or potential source of further information does not mean that the publisher and authors endorse the information or services the organization, website, or product may provide or recommendations it may make. This work is sold with the understanding that the publisher is not engaged in rendering professional services. The advice and strategies contained herein may not be suitable for your situation. You should consult with a specialist where appropriate. Neither the publisher nor authors shall be liable for any loss of profit or any other commercial damages, including but not limited to special, incidental, consequential, or other damages. Further, readers should be aware that websites listed in this work may have changed or disappeared between when this work was written and when it is read.

Library of Congress Cataloging-in-Publication Data

ISBN 9781119794868

Cover image: Refinery I Christian Lagerek | Dreamstime.com
Cover design by Kris Hackerott

Set in size of 11pt and Minion Pro by Manila Typesetting Company, Makati, Philippines

Printed in the USA

10 9 8 7 6 5 4 3 2 1

Companion Web Page

This multi-volume set includes access to its companion web page, from which can be downloaded useful software, spreadsheets, and other value-added products related to the books. To access it, follow the instructions below:

1. Go to https://scrivenerpublishing.com/coker_volume_three/
2. Enter your email in the username field
3. Enter "Refining" in the password field

In Loving Memory of

Dr. Soni O. Oyekan
(A Great Chemical Engineer and Mentor)

The most beautiful spiritual being ever encountered and a privilege to have known. Wishing you the Almighty Father's mercy, grace and blessings, the most wonderful and beautiful Journey in the Luminous Realm of joyful activities in the beyond.

Gratitude to the Elemental and Created beings in higher realms, faithfully fulfilling the Will of the Almighty father.

"God wills that His Laws working in Creation should be quite familiar to man, so that he can adjust himself accordingly, and with their help can complete and fulfill his course through the world more easily and without ignorantly going astray."

Abd-ru-shin
(In the Light of Truth)

The Laws of Creation

The Law of Motion
The Law of the Attraction of Homogeneous Species
The Law of Gravitation
The Law of Reciprocal Action

"What is Truth?"

Pilate (John 18, 38)

"Only the truth is simple."

Sebastian Haffner

"Woe to the people to whom the truth is no longer sacred!"

Friedrich Christoph Schlosser

"Truth does not conform to us, dear son but we have to conform with it."

Matthias Claudius

"Nothing will give safety except truth. Nothing will give peace except the serious search for truth."

Blaise Pascal

“Truth is the summit of being; justice is the application of it to affairs.”

Ralph Waldo Emerson

“The ideals which have lighted my way, and time after time have given me new courage to face life cheerfully, have been Kindness, Beauty and Truth.”

Albert Einstein

“It irritates people that the truth is so simple.”

Johann Wolfgang von Goethe

“Aglow with the Light of the Divine, I surrender my whole attention to the Presence of Truth that guides my path.”

Michael Bernard Beckwith

“Truth means the congruence of a concept with its reality.”

G.W. Friedrich Hegel

“Truth is the revealing gloss of reality.”

Simone Well

“We are the Multi-dimensional Universe becoming aware of Itself. Live in this One Truth – That God is Real As your very Life!”

Michael Bernard Beckwith

“Truth is a torch, but a tremendous one. That is why we hurry past it, shielding our eyes, even terrified of getting burnt.”

Johann Wolfgang von Goethe

“Truth is the spirit’s sun.”

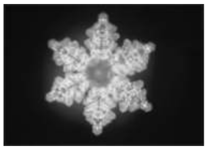
Marquis de Vauvenargues

You will recognise the Truth, and the truth will set you free

John, 8:32

“Truth is the Eternal – Unchangeable! Which never changes in its form, but is as it has been eternally and will ever remain, as it is now. Which can therefore never be subjected to any development either, because it has been perfect from the very beginning. Truth is real, it is ‘being’! Only being is true life. The entire Universe is “supported” by this Truth!”

Abd-ru-shin



Truth

To honour God in all things and to perform everything solely to the glory of God

Abd-ru-shin

(In the Light of Truth)

Awake!

Keep the heart of your thoughts pure, by so doing you will bring peace and be happy.

Love thy neighbour, which means honour him as such!

Therein lies the adamantine command: You must never consciously harm him, either in his body or in his soul, either in his earthly possessions or in his reputation!

He who does not keep this commandment and acts otherwise, serves not God but the darkness, to which he gives himself as a tool!

Honour be to God Who only sows Love! Love also in the The Law of the destruction of the darkness!

Abd-ru-shin

(In the Light of Truth)

Love & Gratitude

Contents

Preface

Petroleum refining is a complex industry that worldwide produces more than $10 billion worth of refined products. Improvements in the design and operation of these facilities can deliver large economic value for refiners. Furthermore, economic, regulatory and environmental concerns impose significant pressure on refiners to provide safe working conditions and at the same time optimize the refining process. Refiners have considered alternative processing units and feedstocks by investing in new technologies.

The United States, Europe and countries elsewhere in the world are embarking on full electrification of automobiles within the next couple of decades. Furthermore, the current pandemic of the coronavirus with lock downs in many countries has restricted the movement of people, less use of aviation fuel and motor gasoline. This has resulted in the barrel of crude being sold at $42.0 per barrel presenting problems to oil producers and refiners. The venture of electrification still poses inherent problems of resolving rechargeable batteries and fuel cells and providing charging stations along various highways and routes. Oil and natural-gas will for the foreseeable future form an important part of everyday life. Their availability has changed the whole economy of the world by providing basic needs for mankind in the form of fuel, petrochemicals and feedstocks for fertilizer plants and energy for the power sector.

Presently, the world economy runs on oil and natural gas, and the processing of these feedstocks for producing fuels, and value-added products has become an essential activity in modern society. The availability of liquefied natural gas (LNG) has enhanced the environment, and recent development in the technology of natural gas to liquids (GTL) has further improved the availability of fuel to transportation and other sectors.

The complex processing of petroleum refining has created a need for environmental, health, and safety management procedures and safe work practices. These procedures are established to ensure compliance with applicable regulations and standards such as hazard communications (PHA, HAZOPS, MoC, and so on), emissions, Waste Management (pollution that includes volatile organic compounds (VOC), carbon monoxide, sulfur oxides (SO_x), nitrogen oxides (NO_x), particulates, ammonia (NH_3), hydrogen sulfide (H_2S), and toxic organic compounds) and waste minimization. These pollutants are often discharged as air emissions, wastewater or solid wastes. Furthermore, concerns over issues such as the depletion of the ozone layer that results in global warming is increasingly having a significant impact on earth's nature and mankind, and carbon dioxide (CO_2) is known to be the major culprit of global warming. Other emissions such as H_2S, NO_x, and SO_x from petroleum refining have adversely impacted the environment, and agencies such as Occupational Safety and Health Administration (OSHA), and Environmental Protection Agency (EPA), Health and Safety Executive (U.K. HSE) have imposed limits on the emissions of these compounds upon refiners.

Flaring has become more complicated and concerns about its efficiency have been increasing and discussed by experts. The OSHA, EPA and HSE have imposed tighter regulations on both safety and emission control, which have resulted in higher levels of involvement in safety, pollution, emissions and so on.

Petroleum refining is one of the important sectors of the world economy, and it's playing a crucial and pivotal role in industrialization, urbanization, and meeting the basic needs of mankind by supplying energy for industrial and domestic transportation, feedstock for petrochemical products as plastics, polymers, agrochemicals, paints, and so on. Globally, it processes more materials than any other industry, and with a projected increase in population to around 8.1 billion by 2025, increasing demand for fuels, electricity and various consumer products made from the petrochemical route is expected via the petroleum and refining process.

Petroleum Refining Design and Applications Handbook, Volume Three, is a continuation of the previous volumes; comprising of four chapters, a revised glossary of petroleum and technical terminology, process data sheets, charts of equilibrium K-values of hydrocarbons, process safety incidents and Excel spreadsheet programs, computer developed programs, UniSim – Design simulation software excises, cases studies and a Conversion Table.

Chapter 18 describes types of mechanical separators, American Petroleum Institute's oil field separators, horizontal gravity settlers and decanters, case studies and chemical safety board (CSB) recommendations. The chapter provides the procedures for sizing two/three-phase oil-gas separator in a horizontal/vertical vessel. Excel spreadsheet programs are used in the examples involving the sizing of these separators, and UniSim Design software is employed in the sizing of a horizontal separator of a wet gas. The chapter provides design procedures for sizing cyclone involving particles as in the fluidized catalytic cracking (FCC) unit of light gas oil (LGO)/heavy gas oil (HGO) from the crude distillation unit (CDU). The chapter further reviews electrical/electrostatic precipitators involving dust or liquid mists, and a process safety case study incident at ExxonMobil refinery in Torrance California. The US chemical safety board (www.csb.gov) investigated the incident in determining the root causes, and provided finding and recommendations. Chapter 19 reviews distillation involving steam distillation, multicomponent distillation, shortcut methods and rigorous using UniSim design software, tower sizing for valve trays, distillation sequencing with columns having more than two products, introducing advanced distillation technology such as cyclic (dividing wall) column and membrane distillation types, troubleshooting, predictive maintenance and controls of distillation columns, heat integration of distillation columns, pinch design approach to inventing a network and heat integration of distillation columns. An important aspect in the chapter is the use of the dividing - wall column in the revamp of the depentanizer, which has found immense use in petroleum refining industry. The chapter further reviews process safety incidents in refinery facilities, and reports by the US Chemical Safety and Investigation board as to how these incidents could best be mitigated and avoided and lessons learned from various root causes. Chapter 20 reviews packed towers involving packing types and materials of construction, service application and provides design procedures for the sizing of a packed tower. The chapter further reviews the guidelines for packing selection and performance, and uses UniSim Design software for simulating multicomponent absorption of wet gas using lean oil. The chapter describes liquid-liquid extraction in the reforming unit of refinery plants. In particular the use of Sulfolane as a solvent to separate the aromatic fractions from other hydrocarbons. The chapter further reviews the solvent selection, process and design parameters required for an extraction unit. UniSim Design software is used in a case study involving the extraction of toluene – n-octane mixture with Sulfolane as a solvent.

Finally, the volume provides a glossary of petroleum and technical terminology, process datasheets, and charts of equilibrium K-values of hydrocarbons, a conversion table, developed Microsoft Excel spreadsheet programs and developed programs including UniSim design software programs that can be readily accessed from the publisher's website using a password.

A. Kayode Coker (www.akctechnology.com)

Acknowledgments

This project is the culmination of five years of research, collating relevant materials from organizations, institutions, companies and publishers, developing Excel spreadsheet programs and computer programs; using Honeywell's UniSim steady state simulation programs and providing the majority of the drawings in the text.

Sincere gratitude to Honeywell Process Solutions for granting permission to incorporate the use of UniSim Design simulation and many other suites of software programs in the book. I express my thanks to Dr. Jamie Barber of Honeywell Process Solutions for his friendship and help over many years of using the UniSim Design simulation software.

To Mr. Ahmed Mutawa formerly of SASREF Co., Saudi Arabia for developing the Conversion Table program for the book.

Many organizations, institutions and companies as Gas Processor Suppliers Association (GPSA), USA, Honeywell Process Solutions, Saudi Aramco Shell Refinery Co. (SASREF), Absoft Corporation, USA, American Institute of Chemical Engineers, The Institution of Chemical Engineers, U.K., Chemical Engineering magazine by Access Intelligence, USA, Hydrocarbon Processing magazine have readily given permission for the use of materials and their release for publication. I greatly acknowledge and express my deepest gratitude to these organizations.

I have been privileged to have met with Phil Carmical, Publisher at Scrivener Publishing Co., some twenty years ago. Phil initiated the well-known Ludwig's project at the time during his tenure at Gulf Publishing Co., and Elsevier, respectively. His suggestions in collaborating on these important works some seven years ago were timely to the engineering community, as I hope that these works will be greatly beneficial to this community world-wide. I'm deeply grateful to Phil for agreeing to collaborate with me, his suggestions and assistance since. I believe that upon completing this aspect of the project that the book will save lives in the refinery industry.

I also wish to express my thanks to the Wiley-Scrivener team: Kris Hackerott- Graphics Designer, Bryan Aubrey – Copy editor, Myrna Ting – Typesetter and her colleagues. I am truly grateful for your professionalism, assistance and help in the production of this volume.

Finally,

Bow down in humility before the Greatness of God,
whose Love is never-ending, and who sends us his
help at all times.
He alone is Life and the Power and the Glory for ever
and ever.

A. Kayode Coker (www.akctechnology.com)

18

Mechanical Separations

Separation is employed in almost oil and gas process, and it is an essential process in achieving the required export specifications. A poorly designed separator will adversely impact downstream unit operations that could result in off-spec oil, gas and water being exported. This can impact costly penalties being applied at the facility. A separator is a unit installed in any process that requires a significant space, reliable safety components and an adequate control system to operate. Space limitations and weight restrictions, e.g., in off-shore platforms mean that the separator must achieve the required separation while minimizing the weight of the vessel and space.

Practically every process operation requires the separation of entrained material or two immiscible phases in a process. This may be either as a step in the purification of one stream, or a principal process operation [1]. These separations may be as follows:

1. liquid particles from vapor or gas
2. liquid particles from immiscible liquid
3. dust or solid particles from vapor or gas
4. solid particles from liquid
5. solid particles from other solids

These operations may sometimes be better known as mist entrainment, decantation, dust collection, filtration, centrifugation, sedimentation, screening, classification, scrubbing, and so on. They often involve handling relatively large quantities of one phase in order to collect or separate the other. Therefore the size of the equipment may become very large. For the sake of space and cost it is important that the equipment be specified and rated to operate as efficiently as possible [2]. This subject will be limited here to the removal or separation of liquid or solid particles from a vapor or gas carrier stream (points 1 and 3 above) or separation of solid particles from a liquid (point 4). Svarosky's review is helpful [3].

Other important separation techniques such as pressure-leaf filtration, centrifugation, rotary drum filtration, and others all require technology very specific to the equipment and cannot be generalized in many instances.

18.1 Particle Size

The particle sizes of liquid and solid dispersoids will vary markedly depending upon the source and nature of the operation generating the particular particles. For design of equipment to reduce or eliminate particles from a fluid stream, it is important either to know from data the range and distribution of particle sizes, or be in a position to intelligently estimate the normal and extreme expectancies. Figures 18.1a and b give a good overall picture of

A. Kayode Coker. Petroleum Refining Design and Applications Handbook Volume 3, (1–248) © 2022 Scrivener Publishing LLC

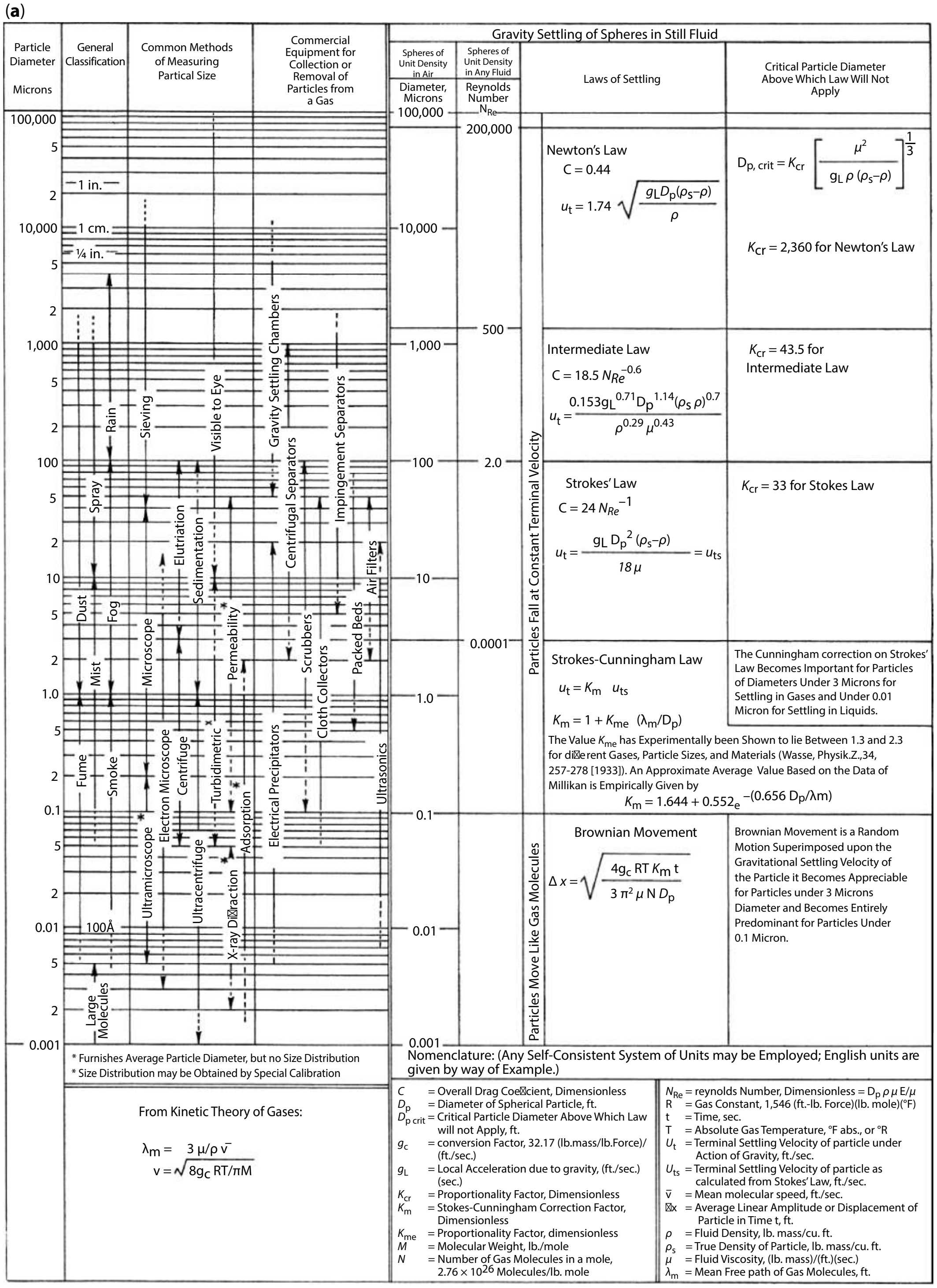
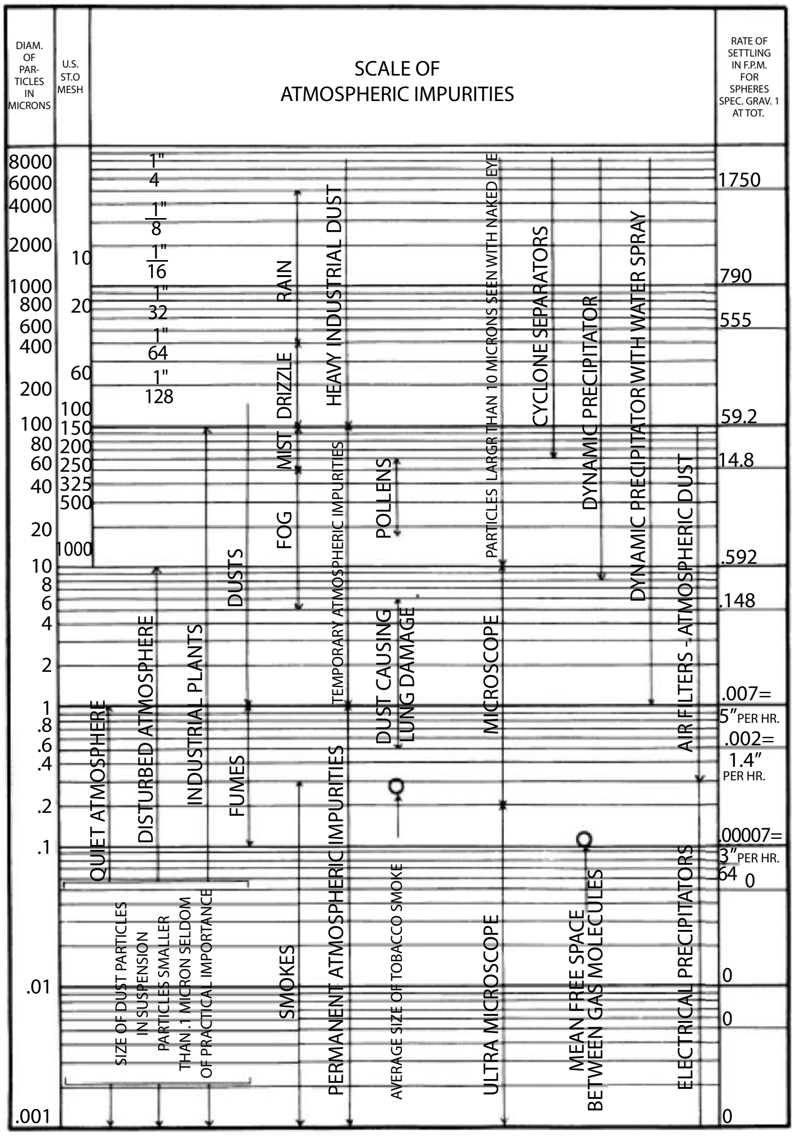


Figure 18.1a Characteristics of dispersed particles (by permission from Perry, J. H., ed., Chemical Engineers Handbook, 3rd ed., 1950 McGraw-Hill Company Inc.).

dimensions as well as the descriptive terminology so important to a good understanding of the magnitude of a given problem. The significant laws governing particle performance in each range is also shown.

Particle sizes are measured in microns (μm). A micron is 1/1000 mm or 1/25,400 in. = 10^{-6} m = 39.37×10^{-6} in. A millimicron, (mμ), is 1/1000 of a micron, or 1/1,000,000 mm. Usually particle size is designated as the average

(b)



IT IS ASSUMED THAT THE PARTICLES ARE OF UNIFORM SPHERICAL SHAPE HAVING SPECIFIC GRAVITY ONE AND THAT THE DUST CONCENTRATION IS 0.6 GRAINS PER 1000 FT^3 OF AIR, THE AVERAGE OF METROPOLITAN DISTRICTS.

Figure 18.1b Size and characteristics of air-borne solids (by permission from Hoffman Handy Engineering Data. Hoffman Air and Filtration Systems, Inc.).

diameter in microns, although some literature reports particle radius. Particle concentration is often expressed as grains/cubic feet of gas volume. One grain (gr) is 1/7000 of a pound.

The mechanism of formation has a controlling influence over the uniformity of particle size and the magnitude of the dimensions. Thus, sprays exhibit a wide particle size distribution, whereas condensed particles such as fumes, mists and fogs are particularly uniform in size. Table 18.1 gives the approximate average particle sizes for dusts and mists, which might be generated around process plants. Figure 18.2 indicates the size ranges for some aerosols, dusts and fumes. Table 18.2 gives typical analysis of a few dusts, and Table 18.3 gives screen and particle size relationships. Table 18.4 gives approximate mean particle size for water spray from a nozzle.

Table 18.1 Sizes of common dusts and mists.

Dust or mist	Average particle diameter, microns (μm)
Human Hair (for comparison)	50–200
Limit of visibility with naked eye	10–40
Dusts	
Atmospheric dust	0.5
Aluminum	2.2
Anthracite Coal Mining	
Breaker air	1.0
Mine air	0.9
Coal Drilling	1.0
Coal Loading	0.8
Rock Drilling	1.0
Alkali fume	1–5
Ammonium Chloride fume	0.05–0.1–1
Catalyst (reformer)	0.5–50
Cement	0.5–40–55
Coal	5–10
Ferro-manganese, or silicon	0.1–1
Foundry air	1.2
Flour mill	15
Fly ash (Boiler flue gas)	0.1–3
Iron (Gray Iron Cupola)	0.1–10
Iron oxide (steel open hearth)	0.5–2
Lime (Lime Kiln)	1–50
Marble cutting	1.5
Pigments	0.2–2
Sandblasting	1.4
Silica	1–10
Smelter	0.1–100
Taconite iron ore (Crushing and Screening)	0.5–100
Talc	10
Talc Milling	1.5
Tobacco smoke	0.2
Zinc oxide fume	0.05

(*Continued*)

Table 18.1 Sizes of common dusts and mists. (*Continued*)

Dust or mist	Average particle diameter, microns (μm)
Zinc (sprayed)	15
Zinc (condensed)	2
Mists	
Atmospheric fog	2–15
Sulfuric Acid	0.5–15

Source: Compiled from References [4–6].

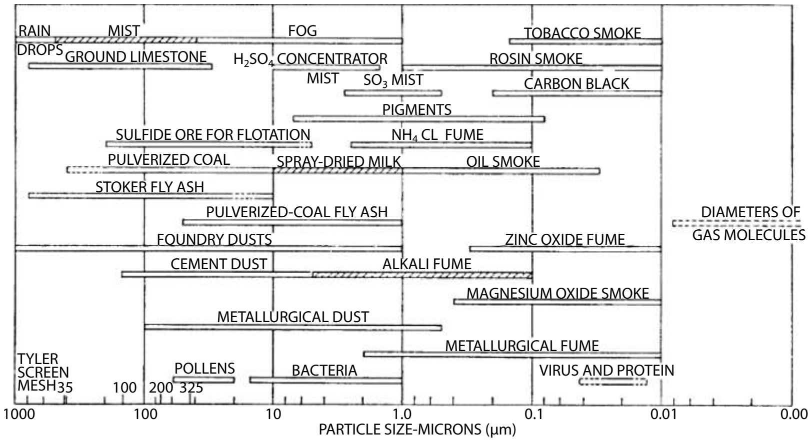


Figure 18.2 Particle size ranges for aerosols, dusts, and fumes (courtesy H.P. Munger, Battle Memorial Institute).

Table 18.2 Typical dust size analysis.*

	DUST			
	Rock	**Cement kiln exhaust**	**Foundry sand**	**Limestone**
Sp. Gravity	2.63	2.76	2.243	2.64
Apparent Wt., lbs./cu ft	61.3	52.0	45.9	72.0
Screen Analysis (% passed)				
100 Mesh	98.8	99.6	91.2	85.6
200 Mesh	92.8	92.2	78.4	76.4
325 Mesh	79.6	80.8	67.6	66.4
400 Mesh	70.8	73.2	64.4	63.2
Elutriation Analysis (% under)				
Terminal Velocity				
320 in./min.	75.8	78.0	64.2	70.5
80 in./min	37.0	61.0	53.9	52.0
20 in./min	17.5	40.8	42.0	33.0
5 in./min	8.9	23.0	32.0	18.0

Source: Compiled from Bulletin No. 1128, American Blower Corp., Detroit, MI.

Table 18.3 Dry particle screen sizes.

W.S. Tyler Screen Scale	Micron (approximate)
80	174
100	146
115	123
150	104
170	89
200	74
250	61
270	53
325	43

Table 18.4 Approximate particle sizes for water spray from liquid full cone spray nozzles.*

Nozzle size (in.)	Operating pressure (psig)	Approximate mean particle size (μm)
½	15	1200
	60	750
¾	10	1600
	40	1000
1	15	1750
	40	1250
1 ½	15	2300
	60	1800
3	10	5300
	30	4300

*Private communication, Spraying Systems Co., Bellwood, III.

18.2 Preliminary Separator Selection

The Sylvan Chart [7] of Figure 18.3 is useful in preliminary equipment selection, although arranged primarily for dust separations; it is applicable in the appropriate parts to liquid separations. Perry [8] presents a somewhat similar chart that is of different form but contains much of the same information as Figures 18.1a and b.

Example 18.1 Basic Separator Type Selection

A suitable collector will be selected for a lime kiln to illustrate the use of the Sylvan Chart (Figure 18.3) [7, 9]. Referring to the chart, the concentration and mean particle size of the material leaving the kiln can vary between 3 and 10 gr/ft^3 with 5–10 μm range of mass mean particle size. Assume an inlet concentration of 7.5 gr/ft^3 and inlet mean particle size of 9 μm. Projection of this point vertically downward to the collection efficiency portion of the chart will indicate that a low resistance cyclone will be less than 50% efficient; a high efficiency centrifugal will be between 60% and 80% efficient, and a wet collector, fabric arrester and electrostatic precipitator will be 97%-plus

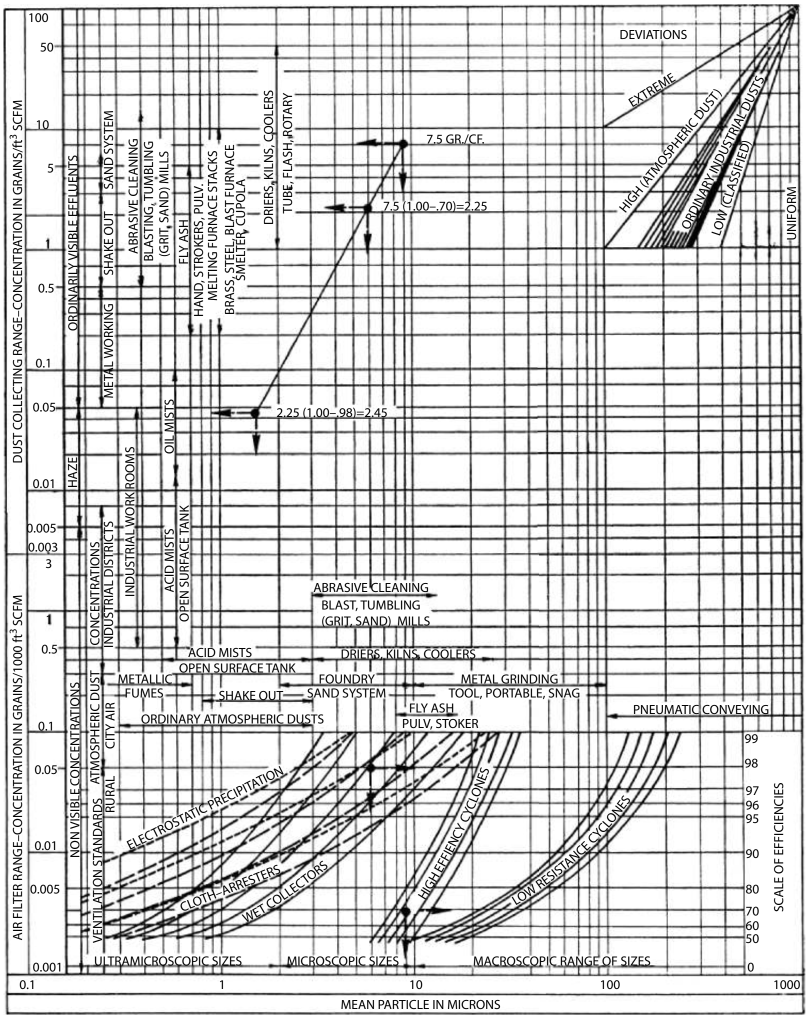


Figure 18.3 Range of particle sizes, concentration, and collector performance (courtesy American Air Filter Co. Inc.).

efficient. The last three collectors are often preceded by a pre-cleaner so a high efficiency centrifugal will be selected. Using the average line of this group, the efficiency will be 70%. Therefore, the effluent from this collector will have a concentration of 7.5(1.00 – 0.70) = 2.25 gr/ft^3.

Draw a line through the initial point with a slope parallel to the lines marked "industrial dust." Where deviation is not known, the average of this group of lines will normally be sufficiently accurate to predict the mean particle size in the collector effluent. Where this line intersects the horizontal line marked 2.25 gr/ft^3, a vertical line through the point will indicate the effluent mean particle size of 6.0 μm.

A projection of this point of collector effluent vertically downward shows that a second-high efficiency centrifugal will be less than 50% efficient. A wet collector, fabric arrester and electrostatic precipitator will be not less than 93% efficient. Selection of a good wet collector will show an efficiency of 98%. The effluent leaving this collector will have a concentration of 2.25 (1.00 – 0.98) = 0.045 gr/ft^3. Using the line initially drawn, at the point where it intersects the line of 0.045 gr/ft^3 will indicate a mean particle size in the effluent of 1.6 μm.

Guide to Dust Separator Applications

Table 18.5 [10] summarizes dry dust particle separators as to general application in industry; Table 18.6, Figures 18.4 and 18.5 [11] compare basic collector characteristics. Figure 18.5 presents a typical summary of dust collection equipment efficiencies which have not changed significantly for many years except for specialized equipment to specialized applications.

Table 18.5 Applications of dust collections in industry.

Operation	Concentration	Particle sizes	Cyclone	High efficiency centrifugal	Wet collector	Fabric arrester	Hi-volt electro-static	See remark no.
Ceramics								
a. Raw product handling	Light	Fine	Rare	Seldom	Frequent	Frequent	No	1
b. Fettling	Light	Fine to medium	Rare	Occasional	Frequent	Frequent	No	2
c. Refractory sizing	Heavy	Coarse	Seldom	Occasional	Frequent	Frequent	No	3
d. Glaze & Vitr. Enamel spray	Moderate	Medium	No	No	Usual	Occasional	No	
Chemicals								
a. Material handling	Light to moderate	Fine to medium	Occasional	Frequent	Frequent	Frequent	Rare	4
b. Crushing grinding	Moderate to heavy	Fine to coarse	Often	Frequent	Frequent	Frequent	No	5
c. Pneumatic conveying	Very heavy	Coarse to fine	Usual	Occasional	Rare	Usual	No	6
d. Roasters, kilns, coolers	Heavy	Coarse to med-coarse	Occasional	Usual	Usual	Rare	Often	7
Coal Mining and Power Plant								
a. Material handling	Moderate	Medium	Rare	Occasional	Frequent	Frequent	No	8
b. Bunker ventilation	Moderate	Fine	Occasional	Frequent	Occasional	Frequent	No	9
c. Dedusting, air cleaning	Heavy	Med-coarse	Frequent	Frequent	Occasional	Often	No	10
d. Drying	Moderate	Fine	Rare	Occasional	Frequent	No	No	11
Fly Ash								
a. Coal burning-chain grate	Light	Fine	No	Rare	No	No	No	12

(*Continued*)

Table 18.5 Applications of dust collections in industry. (*Continued*)

Operation	Concentration	Particle sizes	Cyclone	High efficiency centrifugal	Wet collector	Fabric arrester	Hi-volt electro-static	See remark no.
b. Coal burning-stoker fired	Moderate	Fine to coarse	Rare	Usual	No	No	Rare	
c. Coal burning-pulverized fuel	Heavy	Fine	Rare	Frequent	No	No	Frequent	13
d. Wood burning	Varies	Coarse	Occasional	Occasional	No	No	No	14
Foundry								
a. Shakeout	Light to moderate	Fine	Rare	Rare	Usual	Rare	No	15
b. Sand handling	Moderate	Fine to medium	Rare	Rare	Usual	Rare	No	16
c. Tumbling mills	Heavy	Med-coarse	No	No	Frequent	Frequent	No	17
d. Abrasive cleaning	Moderate to heavy	Fine to medium	No	Occasional	Frequent	Frequent	No	18
Grain Elevator, Flour and Feed Mills								
a. Grain handling	Light	Medium	Usual	Occasional	Rare	Frequent	No	19
b. Grain dryers	Light	Coarse	No	No	No	No	No	20
c. Flour dust	Moderate	Medium	Usual	Often	Occasional	Frequent	No	21
d. Feed mill	Moderate	Medium	Usual	Often	Occasional	Frequent	No	22
Metal Melting								
a. Steel blast furnace	Heavy	Varied	Frequent	Rare	Frequent	No	Frequent	23
b. Steel open hearth	Moderate	Fine to coarse	No	No	Doubtful	Possible	Probable	24
c. Steel electric furnace	Light	Fine	No	No	Considerable	Frequent	Rare	25
d. Ferrous cupola	Moderate	Varied	Rare	Rare	Frequent	Occasional	Occasional	26

(*Continued*)

Table 18.5 Applications of dust collections in industry. (*Continued*)

Operation	Concentration	Particle sizes	Cyclone	High efficiency centrifugal	Wet collector	Fabric arrester	Hi-volt electro-static	See remark no.
e. Non-ferrous reverberatory	Varied	Fine	No	No	Rare	?	?	27
f. Non-ferrous crucible	Light	Fine	No	No	Rare	Occasional	?	28
Metal Mining and Rock Products								
a. Material handling	Moderate	Fine to medium	Rare	Occasional	Usual	Considerable	?	29
b. Dryers, kilns	Moderate	Med-coarse	Frequent	Frequent	Frequent	Rare	Occasional	30
c. Cement rock dryer	Moderate	Fine to medium	Rare	Frequent	Occasional	No	Occasional	31
d. Cement kiln	Heavy	Fine to medium	Rare	Frequent	Rare	No	Considerable	32
e. Cement grinding	Moderate	Fine	Rare	Rare	No	Frequent	Rare	33
f. Cement clinker cooler	Moderate	Coarse	Occasional	Occasional	?	?	?	34
Metal Working								
a. Production grinding, scratch brushing, abrasive cut off	Light	Coarse	Frequent	Frequent	Considerable	Considerable	No	35
b. Portable and swing frame	Light	Medium	Rare	Frequent	Frequent	Considerable	No	
c. Buffing	Light	Varied	Frequent	Rare	Frequent	Rare	No	36
d. Tool room	Light	Fine	Frequent	Frequent	Frequent	Frequent	No	37
e. Cast iron machining	Moderate	Varied	Rare	Frequent	Considerable	Considerable	No	38
Pharmaceutical and Food Products								
a. Mixers, grinders, weighing, blending, bagging, packaging	Light	Medium	Rare	Frequent	Frequent	Frequent	?	39

(*Continued*)

Table 18.5 Applications of dust collections in industry. (*Continued*)

Operation	Concentration	Particle sizes	Cyclone	High efficiency centrifugal	Wet collector	Fabric arrester	Hi-volt electro-static	See remark no.
b. Coating pans	Varied	Fine to medium	Rare	Rare	Frequent	Frequent	No	40
Plastics								
a. Raw material processing	(See comments under Chemicals)							41
b. Plastic finishing	Light to moderate	Varied	Frequent	Frequent	Frequent	Frequent	No	42
Rubber Products								
a. Mixers	Moderate	Fine	No	No	Frequent	Usual	No	43
b. Batchout rolls	Light	Fine	No	No	Usual	Frequent	No	44
c. Talc dusting and dedusting	Moderate	Medium	No	No	Frequent	Usual	No	45
d. Grinding	Moderate	Coarse	Often	Often	Frequent	Often	No	46
Woodworking								
a. Wood working machines	Moderate	Varied	Usual	Occasional	Rare	Frequent	No	47
b. Standing	Moderate	Fine	Frequent	Occasional	Occasional	Frequent	No	48
c. Waste conveying, hogs	Heavy	Varied	Usual	Rare	Occasional	Occasional	No	49

By permission, John M. Kane, Plant Engineering, Nov. (1954).

Table 18.5 Applications of dust collections in industry. (*Continued*)

Remarks referred to in Table 18.5			
1.	Dust released from bin filling, conveying, weighing, mixing, pressing, forming. Refractory products, dry pan, and screening operations more severe.	27.	Zinc oxide loading heavy during zinc additions. Stack temperatures high.
2.	Operations found in vitreous enameling, wall and floor tile, pottery.	28.	Zinc oxide plume can be troublesome in certain plant locations.
3.	Grinding wheel or abrasive cut off operation. Dust abrasive.	29.	Crushing, screening, conveying, storing involved. Wet ores often introduce water vapor in exhaust air stream.
4.	Operations include conveying, elevating, mixing, screening, weighing, packaging. Category covers so many different materials that recommendation will vary widely.	30.	Dry centrifugals used as primary collectors, followed by final cleaner.
5.	Cyclone and high efficiency centrifugals often act as primary collectors followed by fabric or wet type.	31.	Collection equipment installed primarily to prevent public nuisance.
6.	Usual set up uses cyclone as product collector followed by fabric arrester for high over-all collection efficiency.	32.	Collection usually permits salvage of material and also reduce nuisance from settled dust in plant area.
7.	Dust concentration determines need for dry centrifugal; plant location, product value determines need for final collectors. High temperatures are usual and corrosive gases not unusual.	33.	Salvage value of collected material high. Same equipment used on raw grinding before calcining.
8.	Conveying, screening, crushing, unloading.	34.	Coarse abrasive particles readily removed in primary collector types.
9.	Remote from other dust producing points. Separate collector usually.		
10.	Heavy loading suggests final high efficiency collector for all except very remote locations.	35.	Roof discoloration, deposition on autos can occur with cyclones and less frequently with dry centrifugal. Heavy duty air filters sometimes used as final cleaners.
11.	Difficult problem but collectors will be used more frequently with air pollution emphasis.	36.	Linty particles and sticky buffing compounds can cause trouble in high efficiency centrifugals and fabric arresters. Fire hazard is also often present.
12.	Public nuisance from boiler-down indicates collectors are needed.	37.	Unit collectors extensively used, especially for isolated machine tools.
13.	Higher efficiency of electrostatic indicated for large installations especially in residential locations. Often used in conjunction with dry centrifugal.	38.	Dust ranges from chips to fine floats including graphitic carbon.
14.	Public nuisance from sealed wood char indicates collectors are needed.	39.	Materials involved vary widely. Collector selection may depend on salvage value, toxicity, sanitation yardsticks.

(*Continued*)

Table 18.5 Applications of dust collections in industry. (*Continued*)

Remarks referred to in Table 18.5			
15.	Hot gases and steam usually involved.	40.	Controlled temperature and humidity of supply air to coating pans makes recirculation from coating pans desirable.
16.	Steam from hot sand, adhesive clay bond involved.	41.	Manufacture of plastic compounds involve operations allied to many in chemical field and vary with the basic process employed.
17.	Concentration very heavy at start of cycle.	42.	Operations are similar to woodworking and collection selection involves similar considerations. See Item 13.
18.	Heaviest load from airless blasting due to higher cleaning speed. Abrasive shattering greater with sand than with grit or shot. Amounts removed greater with sand castings, less with forging scale removal, least when welding scale is removed.	43.	Concentration is heavy during feed operation. Carbon black and other fie additions make collection and dust free disposal difficult.
19.	Operations such as car unloading, conveying, weighing, storing.	44.	Often no collection equipment is used where dispersion from exhaust stack is good and stack location favorable.
20.	Collection equipment expensive but public nuisance complaints becoming more frequent.	45.	Salvage of collected material often dictates type of high efficiency collector.
21.	In addition to grain handling, cleaning, rolls, sifters, purifiers, conveyors, as well as storing, packaging operations are involved.	46.	Fire hazard from some operations must be considered.
22.	In addition to grain handling, bins, hammer mills, mixers, feeders, conveyors, bagging operations and control.	47.	Bulky material. Storage for collected material is considerable, bridging from splitters and chips can be a problem.
23.	Primary dry trap and wet scrubbing usual. Electrostatic is added where maximum cleaning required.	48.	Production sanding produces heavy concentration of particles too fine to be effectively caught by cyclones or dry centrifugals.
24.	Cleaning equipment seldom installed in past. Air pollution emphasis indicates collector use will be more frequent in future.	49.	Primary collector invariably indicated with concentration and partial size range involved, wet or fabric collectors when used are employed as final collectors.
25.	Where visible plume objectionable from air pollution standards, use of fabric arresters with greater frequent seems probable.		
26.	Most cupolas still have no collectors but air pollution and public nuisance emphasis is creating greater interest in control equipment.		

Table 18.6 Comparison of some important dust collector characteristics.*

Type	Higher efficiency range on particles greater than mean size in microns	Pressure loss, (inches of water)	Water (gal. per 1000 cfm)	Space	Sensitivity to cfm change			
					Pressure	Efficiency	Humid air influence	Max. temp., (°F) (standard construction)
Electro-Static	0.25	½	–	Large	Negligible	Yes	Improves efficiency	500
Fabric: Conventional	0.4	3–6	–	Large	As cfm	Negligible	May make reconditioning difficult	180
Reverse Jet	0.25	3–8	–	Moderate	As cfm	Negligible		200
Wet:								
Packed Tower	1–5	1½–3½	5–10	Large	As cfm	Yes		
Wet Centrifugal	1–5	2½–6	3–5	Moderate	As (cfm)^2	Yes		
Wet Dynamic	1–2	Note 1	½ to 1	Small	Note 1	No	None	Unlimited
Orifice Types	1–5	2½–6	10–40	Small	As cfm	Varies with		
					or less	design		
Higher Efficiency:								
Nozzle	0.5–5	2–4	5–10	Moderate	As (cfm)^2	Slightly to	None	Note 2
Venturi	0.5–2	12–20	–	Small		Moderately		Unlimited
Dry Centrifugal:								
Low Pressure Cycle	20–40	¾–1½	–	Large	As (cfm)^2	Yes	Many cause	750
High Eff. Centrif.	10–30	3–6	–	Moderate	As (cfm)^2	Yes	Condensation	750
Dry Dynamic	10–20	Note 1	–	Small	Note 1	No	and plugging	750
Louver	15–60	1–3	–	Small	As (cfm)^2	Moderately		750

Note 1: A function of the mechanical efficiency of these combined exhauster and dust collectors.

Note 2: Precooling of high temperature gases will be necessary to prevent rapid evaporation of fine droplets.

Source: *By permission from John M. Kane, "Operation, Application and Effectiveness of Dust Collection Equipment," Heating and Ventilating, Aug. 1952, Ref. (10).

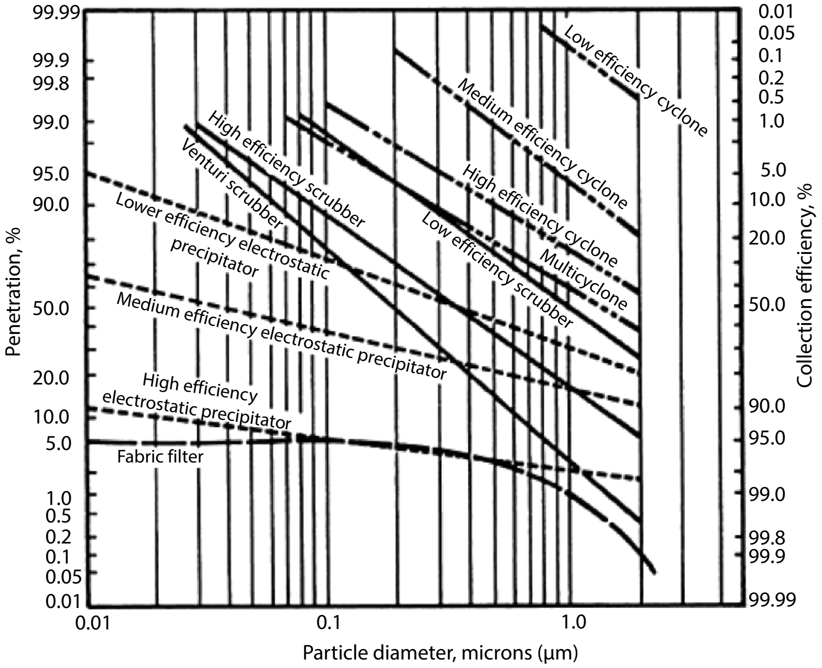


Figure 18.4 Comparison chart showing ranges of performance of several collection/control devices in air streams (by permission from Vandergrift *et al.* Chemical Engineering, Deskbook Issue, Jan 18, p. 109, 1973).

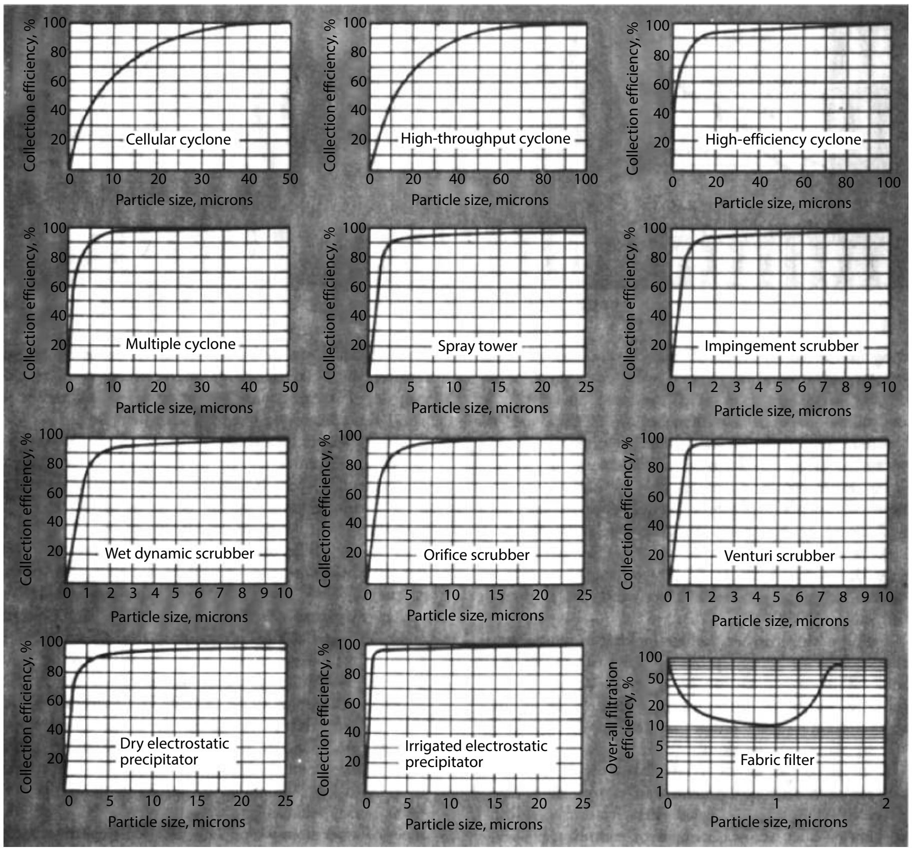


Figure 18.5 Efficiency curves for various types of dust collection equipment as of 1969. Only marginal improvements have been made since then (by permission from Sargent, C. D., Chemical Engineering, p. 130, Jan 27, 1969).

Table 18.7 General applications of liquid particle separators.

				Collector types			
Operation	Concentration	Particle size	Gravity	Impingement	Cyclone	Scrubbers	Electrical
Pipeline entrained liquid	Light	Fine to coarse	No	Frequent	Yes	Occasional	Few
Compressor discharge liquid	Light	Fine	No	Frequent	Occasional	Occasional	Rare
Compressor oil haze	Vary light	Very fine	No	Frequent	Frequent	Frequent	Occasional
Flashing liquid	Light to mod.	Fine to medium	No	Frequent	Frequent	Occasional	Rare
Boiling or bubbling	Light to heavy	Fine to coarse	Occasional	Frequent	Frequent	Occasional	Rare
Spraying	Light to heavy	Fine to coarse	No	Frequent	Frequent	Rare	Rare
Corrosive liquid particles	Light to heavy	Fine to heavy	Occasional	Frequent	Occasional	Frequent	Rare
Liquid plus solid particles	Light to heavy	Medium	Occasional	Occasional	Frequent	Frequent	Occasional

Guide to Liquid–Solid Particle Separators

Table 18.7 summarizes liquid particle separators as to the general process-type application.

18.3 Gravity Settlers

The use of these settlers is not usually practical for most situations. The diameters or cross-section areas become too large for the handling of anything but the very smallest of flowing vapor streams. In general, gravity settlers of open box or tank design are not economical for particles smaller than 325 mesh or 43 μm [8]. They are much more practical for solids or dusts, although even for these situations the flow quantities must be small if the sizes are not to become excessive. With unusually heavy and/or large particles the gravity separator can be used to advantage. The motion of particle and fluid are considered relative, and the handling of the relations is affected only by conditions of turbulence, eddy currents, and so on.

The basic principles used to achieve physical separation of gas and liquids or solids are momentum, gravity settling and coalescing. Any separator may employ one or more of these principles, but the fluid phases must be immiscible and have different densities (in the case of oil, gas and water) for separation to occur. More dense fluids sink to the bottom, while less dense fluids rise to the top. In the case of liquid emulsions, heating may be used to encourage the separation.

Considering two-phases that are mixed together but are immiscible from one another, one phase will form droplets (referred to as the dispersed phase) within the other phase (referred to as the continuous phase). When there are droplets of water in oil then the water droplets will have forces acting on them in the vertical direction; gravity and drag (Figure 18.6). These forces will be unbalanced, which will cause the droplet to move. Initially, the gravitational force is greater than the drag force, and the droplet will experience acceleration downwards. As the vertical velocity

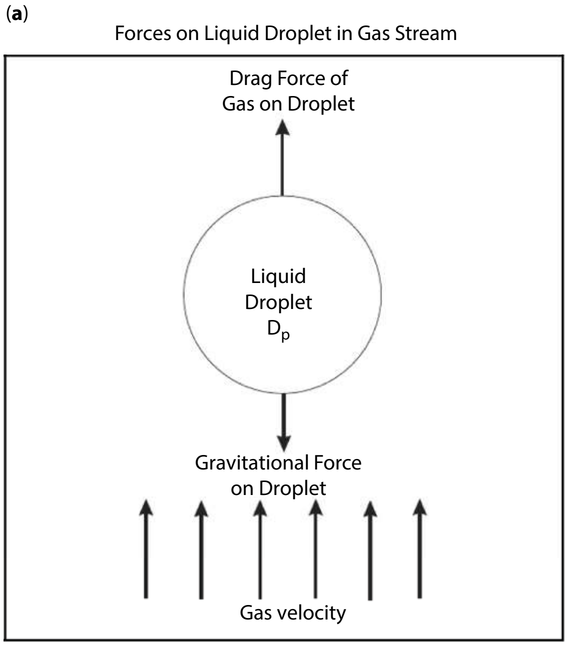


Figure 18.6a Schematic of the forces acting on a liquid droplet in gas phase.

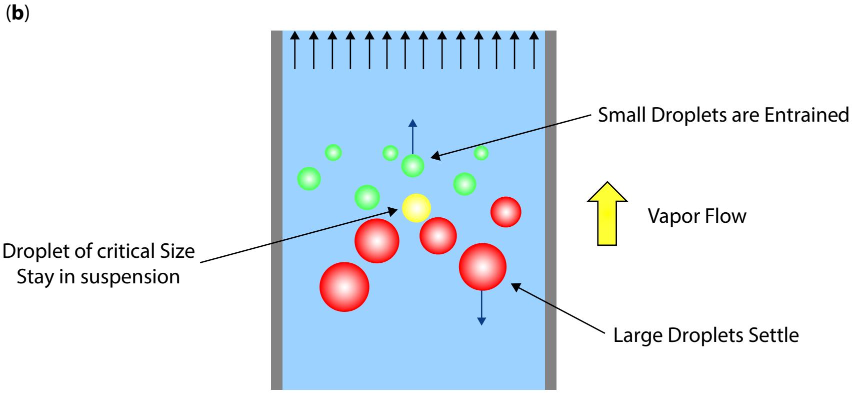


Figure 18.6b Gravity settling theory.

of the droplet increases, the drag force also increases. This will continue until a terminal velocity u_t is reached and the drag and gravitational forces acting on the droplet are equal and no further acceleration is experienced.

At equilibrium, these forces are equal

$$F_D = F_G \tag{18.1}$$

The drag force is expressed as:

$$F_D = C_D\, A_p \frac{\rho_G v^2}{2g_c} \tag{18.2}$$

The droplet projected area, A_p is:

$$A_p = \left(\frac{\pi D_p^2}{4}\right) \tag{18.3}$$

Substituting Eq. (18.3) into Eq. (18.2) gives:

$$F_D = C_D \frac{\pi D_p^2}{4} \frac{\rho_G v^2}{2g_c} \tag{18.4A}$$

The gravitational force F_G on a sphere from Archimedes' principle is:

$$F_G = (\rho_L - \rho_G) V_p \frac{g}{g_c} \tag{18.4B}$$

The volume of spherical droplet, V_p is:

$$V_p = \left(\frac{\pi D_p^3}{6}\right) \tag{18.5}$$

Substituting Eq. (18.5) into Eq. (18.4) gives:

$$F_G = (\rho_L - \rho_G) \frac{\pi D_p^3}{6} \frac{g}{g_c} \tag{18.6}$$

Equating Eqs. (18.4) and (18.6) and solving for the gas maximum velocity gives:

$$C_D \frac{\pi D_p^2}{4} \rho_G \frac{v_{G\max}^2}{2g_c} = (\rho_L - \rho_G) \frac{\pi D_p^3}{6} \frac{g}{g_c} \tag{18.7}$$

$$v_{G\max} = \sqrt{\left(\frac{4gD_p}{3C_D}\right)\left(\frac{\rho_L - \rho_G}{\rho_G}\right)} \tag{18.8}$$

where

$$K_s = \sqrt{\left(\frac{4gD_p}{3C_D}\right)} \tag{18.9}$$

Therefore, the maximum gas velocity which prevents entrainment of liquid is:

$$v_{G\max} = K_s \sqrt{\left(\frac{\rho_L - \rho_G}{\rho_G}\right)} \tag{18.10}$$

Eq. 18.10 is known to as Souder–Brown equation and K_s is the design parameter.

where
A_p = projected area of the droplet, ft^2
C_D = Drag coefficient
g = Acceleration due to gravity, 32.2 ft/s^2
g_c = Conversion factor $\left(32.174 \frac{lb_m}{lb_f} \cdot \frac{ft}{s^2}\right)$
v = Gas velocity, ft/s
V_p = Volume of droplet, ft^3
ρ_G = Gas density, lb/ft^3
ρ_L = Liquid density, lb/ft^3

Once the maximum velocity, v_{Gmax} gas velocity through the vessel is determined, the required minimum cross-sectional area of the vessel for gas flow is:

$$A_{G\min} = \left(\frac{\pi D_{\min}^2}{4}\right) F_G = \frac{q_a}{v_{G\max}} \tag{18.11}$$

Solving for the minimum vessel diameter, D_{min} gives:

$$D_{\min} = \sqrt{\frac{(4/\pi) q_a}{(F_G\, v_{G\max})}} \tag{18.12}$$

where
F_G = Fraction of cross-sectional area available for gas flow (F_G = 1 for vertical separators and it is a function of liquid height for horizontal separators).
q_a = Gas flow rate at the actual flowing conditions

18.4 Terminal Velocity

When a particle falls under the influence of gravity it will accelerate until the frictional drag in the fluid balances the gravitational forces. At this point it will continue to fall at constant velocity. This is the terminal velocity or free-settling velocity. The general formulae for any shape particle are [5]:

$$u_t = \sqrt{\frac{2 g_L m_p (\rho_s - \rho)}{\rho \rho_s A_p C}} \tag{18.13}$$

where
A_p = area if particle projected on plane normal to direction of flow or motion, ft^2.
C = overall drag coefficient, dimensionless
g_L = acceleration due to gravity, 32.2 ft/s^2
m_p = mass of particle, lb_m
ρ = fluid density, lb/ft^3
ρ_s = particle density, lb/ft^3
u_t = terminal settling velocity determined by Stokes' Law of particle under action of gravity, ft/s

For spheres:

$$u_t = \sqrt{\frac{4g_L D_p(\rho_s - \rho)}{3\rho C}} \tag{18.14}$$

$$K_s = \sqrt{\frac{4g_L\, D_p}{3C}} \tag{18.15}$$

Substituting Eq. 18.15 into Eq. 18.14 gives:

$$u_t = K_s\sqrt{\frac{(\rho_s - \rho)}{\rho}} \tag{18.16}$$

where
D_p = diameter of particle, ft

(a) Spherical particles between 1500 and 100,000 μm; Newton's Law:

$$u_t = 1.74\sqrt{\frac{g_L D_p(\rho_s - \rho)}{\rho}} \tag{18.17}$$

C = 0.445 average drag coefficient

(b) Spherical particles between 100 and 1500 μm (with $C = 18.5 N_{Re}^{-0.6}$ (see Figure 18.1)) [5].

$$u_t = \frac{0.153 g_L^{0.71} D_p^{1.14}(\rho_s - \rho)^{0.7}}{\rho^{0.29}\mu^{0.43}} \tag{18.18}$$

where
μ = fluid viscosity, cP (1cP = 6.72 × 10^{-4} lb/ft.s).

The drag coefficient, C_D (C) is a function of Reynolds number, $N_{Re} = \frac{\rho_G\, v\, D_p}{\mu_G}$. For Stokes' law

$$C_D = \frac{24}{N_{Re}} = \frac{24\mu_G}{\rho_G\, v\, D_p} \text{ for } N_{Re} < 2 \tag{18.19}$$

(c) For spherical particles between 3 and 100 μm and Reynolds numbers between 0.0001 and 2.0, Stokes Law:

$C = 24 N_{Re}^{-1}$

$F_d = 3\pi\mu u D_p/g_L$

where
u = relative velocity particle and main body of fluid, ft/s.

Substituting C_D from Eq. 18.19 into Eq. 18.8 gives liquid droplet terminal velocity, u_t, in the gas phase based on the Stokes' law.

$$u_t = g_L D_p^2 \frac{(\rho_s - \rho)}{18\mu} \tag{18.20}$$

For particles smaller than 0.1 μm the random Brownian motion is greater than the motion due to gravitational settling. Therefore, the above relations based on Stokes' Law will not hold.

(d) Spherical particles between 0.1 and 3 μm, by Stokes–Cunningham Law [12]:

$$u_t = K_m u_{ts} \tag{18.21}$$

where

u_t = terminal settling velocity determined by Stokes' law, ft/s.
K_m = Stokes–Cunningham correction factor, dimensionless.
u_{ts} = terminal settling velocity, as calculated from Stokes' law, ft/s.

$$K_m = 1 + K_{me}\left(\lambda_m / D_p\right) \tag{18.22}$$

$$K_{me} = 1.64 + 0.552 e^{-\left(0.656 D_p/\lambda_m\right)} \tag{18.23}$$

This represents a correction on Stokes law and is significant for 3 μm and smaller particles in gases and 0.01 μm and smaller particles in liquids. Table 18.8 gives values of K_m.

When two free settling particles of different dimensions, D'_{p1} and D'_{p2} and different densities, ρ_{p1} and ρ_{p2}, fall through a fluid of density, ρ_f, they will attain equal velocities when:

$$\frac{D'_{p1}}{D'_{p2}} = \left(\frac{\rho_{p2} - \rho_f}{\rho_{p1} - \rho_f}\right)^n \tag{18.24}$$

where n = 1 in eddy resistance zone (more turbulent) and n = 0.5 in streamline fall.

The drag coefficient has been found to be a function of the shape of the particle and the Reynolds number of the flowing gas. In Eqs. 18.13 and 18.14, the particle shape is considered to be a solid, rigid sphere. The Reynolds number is defined as

Imperial Units

$$Re = 1488 \frac{D_p u_t \rho_g}{\mu} \tag{18.25}$$

where

D_p = droplet diameter, ft.
u_t = terminal velocity for the particle size D_p to drop or settle out of gas, ft/s
ρ_g = gas phase density, lb/ft^3
μ = viscosity of continuous phase, cP

Table 18.8 Values of K_m for air at atmospheric pressure [12].

Particles diameter, D_p (μm)	**70°F**	**212°F**	**500°F**
0.1	2.8	3.61	5.14
0.25	1.682	1.952	2.528
0.5	1.325	1.446	1.711
1.0	1.160	1.217	1.338
2.5	1.064	1.087	1.133
5.0	1.032	1.043	1.067
10.0	1.016	1.022	1.033

m = Exponent given by equations in Reynolds number table below
V_t = Settling velocity for single spherical particle, ft/s and m/s (terminal)
V_{ts} = Settling velocity for hindered uniform spherical particle, ft/s or m/s (terminal)
c = Volume fraction solids
K = Constant given by equation above
N_{Re} = Reynolds number, $D_p V_t \rho_f/\mu$

Values of m	**N_{Re}**
4.65	<0.5
$4.375 (N_{Re})^{-0.0875}$	$0.5 \le N_{Re} \le 1{,}300$
2.33	$N_{Re} > 1{,}300$

SI Units

$$\mathrm{Re} = 1000 \frac{D_p u_t \rho_g}{\mu} \tag{18.26}$$

where

D_p = droplet diameter, m.
u_t = terminal velocity for the particle size D_p to drop or settle out of gas, m/s
ρ_g = gas phase density, kg/m^3
μ = viscosity of continuous phase, mPa.s

Figure 18.7 shows the relationship between drag coefficient and particle Reynolds number for spherical particles. In this form, a trial and error solution is required since both particle size, D_p and terminal velocity, u_t are involved. Figure 18.8 presents values of the product of drag coefficient (C′) times the Reynolds number squared, as this technique eliminates velocity from the expression. The abscissa of Figure 18.8 is given by:

$$C'(\mathrm{Re})^2 = \frac{(0.95)(10^8)\rho_g D_p^3(\rho_l - \rho_g)}{\mu^2} \tag{18.27}$$

C′ or C = drag coefficient of particle, dimensionless (Figure 18.7)
ρ_1 = liquid phase density, droplet or particle, lb/ft^3

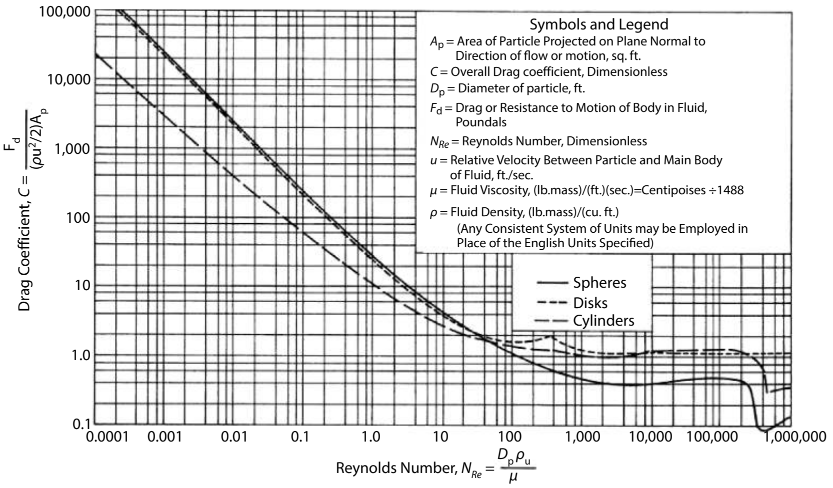


Figure 18.7 Drag coefficients for spheres, disks, and cylinders in any fluid (by permission from Perry, J. H., ed., Chemical Engineering Handbook, 3rd ed., McGraw-Hill Company 1950).

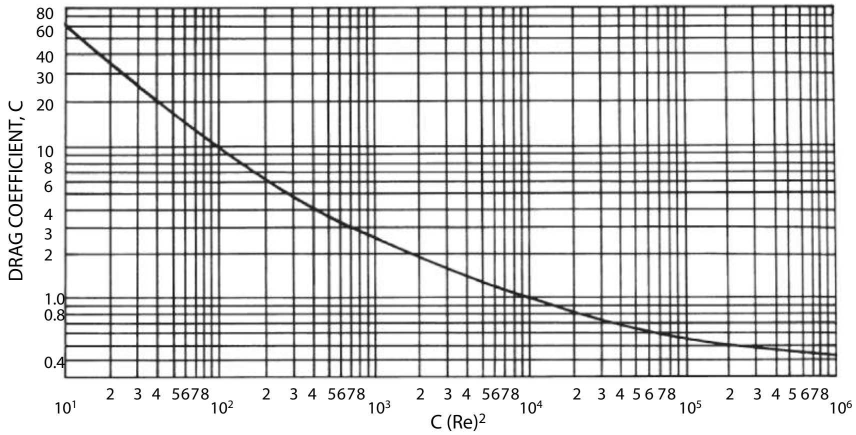


Figure 18.8 Drag coefficient of rigid spheres (source: Gas Processors Suppliers Association (GPSA) Engineering Data Book 12th ed.).

SI Units

$$C'(Re)^2 = \frac{(1.31)(10^7)\rho_g D_p^3(\rho_l - \rho_g)}{\mu^2} \tag{18.28}$$

where

C′ or C = drag coefficient of particle, dimensionless (Figure 18.7)

ρ_1 = liquid phase density, droplet or particle, kg/m^3

18.5 Alternate Terminal Velocity Calculation

In contrast to individual particles settling in a very dilute solution/fluid, is the case of sedimentation where particles must settle in more concentrated environment, and hence, particles influence adjacent particles. This is often termed hindered settling [8, 13]. Depending upon the particles concentration, the hindered terminal settling velocity will generally be somewhat lower than for the terminal settling velocity of a single desired particle in the same medium.

According to Carpenter, K is a dimensionless number that establishes the regime of settling class, reference to the settling laws [13]:

$$K = 34.8\, D'_p \left[\frac{\rho_f (\rho_p - \rho_f)}{\mu^2} \right]^{1/3} \tag{18.29}$$

If, K < 3.3, then Stokes' law applies. If 3.3 ≤ K ≤ 43.6, the intermediate law applies, and if K > 43.6, Newton's law applies. If K > 2360, the equations should not be used. The corresponding b_1 and a values for these K values are tabulated below.

Law	**Range**	$\mathbf{b_1}$	**a**
Stokes'	K < 3.3	24.0	1.0
Intermediate	3.3 ≤ K ≤ 43.6	18.5	0.6
Newton's	K > 43.6	0.44	0

The terminal settling velocity for single spheres can be determined using the contrasts for the flow regime.

$$V_t = \left[\frac{4 a_e D'^{(1+n)}_p (\rho_p - \rho_f)}{3 b_1 \mu^n \rho_f^{(1-n)}} \right]^{1/(2-n)}, \text{ft/s.} \tag{18.30}$$

For hindered particle settling in a "more crowded" environment, using spherical particles of uniform size:

$$V_{ts} = V_t(1 - c)^m, \text{ ft/s} \tag{18.31}$$

where

c = volume fraction solids
D'_p = diameter of particle, in. or mm
a_e = acceleration due to gravity, 32.2 ft/s^2 or 9.8 m/s^2
ρ_p = density of particles, lb/ft^3 or kg/m^3
ρ_f = density of fluid, lb/ft^3 or kg/m^3
μ = viscosity of fluid, cP
b_1 = constant given above
K = constant given by equation above
m = exponent given by equations in Reynolds number table below.
n = constant given in text.
V_t = settling velocity for single spherical particle, ft/s and m/s (terminal).
V_{ts} = settling velocity for hindered uniform spherical particle, ft/s or m/s (terminal).

Referring the above to other than uniform spherical particles does not create a significant loss in accuracy for industrial applications. For higher concentration, the values of V_{ts} are lower than V_t. In large particles in small vessels, the wall effect can become significant [8].

For a single particle, D_p can be taken as 2 (hydraulic radius), and the Sauter mean diameter for hindered particles.

Reynolds number N_{Re}:

$$N_{Re} = \frac{D_p V_t \rho_f}{\mu}, \text{dimensionless} \tag{18.32}$$

Values of m	N_{Re}
4.65	<0.5
$4.375(N_{Re})^{-0.0875}$	$0.5 \le N_{Re} \le 1300$
2.33	$N_{Re} > 1300$

Example 18.2 Hindered Settling Velocities

Using the example of Carpenter [13]:

ρ_f = fluid density = 0.08 lb/ft³
μ = viscosity = 0.02 cP
ρ_p = 500 lb/ft³
D'_p = particle diameter, in. = 0.01
c = volume fraction solids, 0.1

Solving equation for K, for unhindered particle:

$$K = 34.81(0.01)\left[\frac{0.08(500-0.08)}{(0.02)^2}\right]^{1/3}$$
$$K = 16.16$$

Then, for K = 16.16 (intermediate range), b = 18.5; n = 0.6

Solving Eq. (18.30) for settling velocity, V_t:

$$V_t = \left[\frac{4(32.2)(0.01)^{(1+0.6)}(500-0.08)}{3(18.5)(0.02)^{0.6}(0.08)^{(1-0.6)}}\right]^{1/(2-0.6)}$$
$$V_t = 8.81 \text{ ft/s}$$

$$\text{Reynolds number}, N_{Re} = \frac{D_p V_t \rho_f}{\mu}$$
$$= \left(\left[\frac{0.01}{(12)}\right]\frac{(8.81)(0.08)}{(0.02)(6.72\times10^{-4})}\right)\left(\text{in.}\frac{\text{ft.}}{\text{in}}\cdot\frac{\text{ft}}{\text{s}}\cdot\frac{\text{lb}}{\text{ft}^3}\cdot\frac{1}{\frac{\text{lb}}{\text{ft.s}}}\right)$$

$$N_{Re} = 43.70$$
$$\mu = (\text{cP})(6.72\times10^{-4}), \text{lb/ft s}$$

Then, $m = 4.375(N_{Re})^{-0.0875} = 4.375(43.70)^{-0.0875} = 3.1437$

For 0.1 volume fraction solids for hindered settling velocity:

$$\begin{aligned} V_{ts} &= V_t(1 - c)^m \\ &= 8.81(1 - 0.1)^{3.1437} \\ &= 6.33 \text{ ft/s} \end{aligned}$$

(e) ***Particles under 0.1 μm***: Brownian movement becomes appreciable for particles under 3 μm and predominates when the particle size reaches 0.1 μm [5]. This motion usually has little effect in the average industrial process settling system except for the very fine fogs and dusts. However, this does not mean that problems are not present in special applications.

Figure 18.1a gives the limiting or critical diameter above which the particular settling law is not applicable. Figure 18.9 gives terminal velocities for solids particles falling in standard air (70°F and 14.7 psia), and Figure 18.10 gives

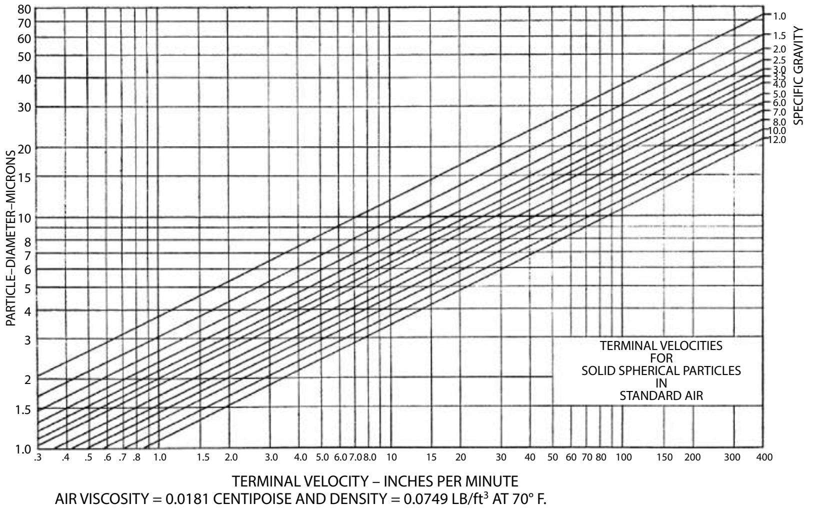


Figure 18.9 Terminal velocities of solid particles in standard air (courtesy of American Blower Div., American Radiator and Sanitary Corporation).

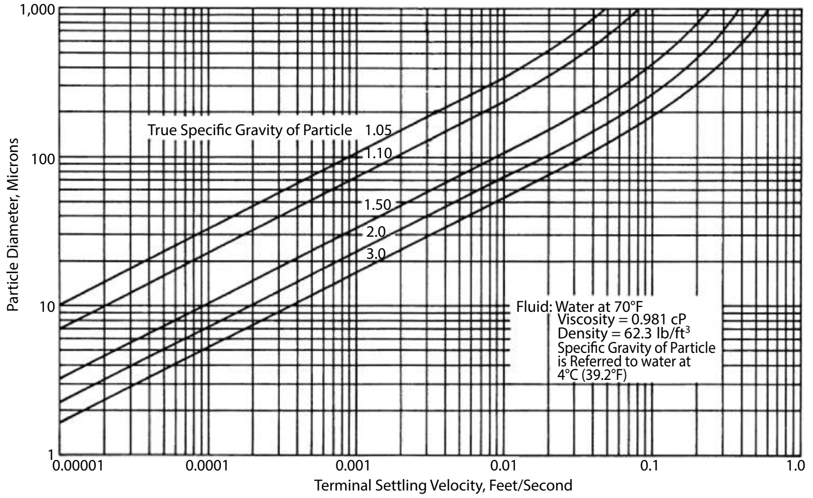
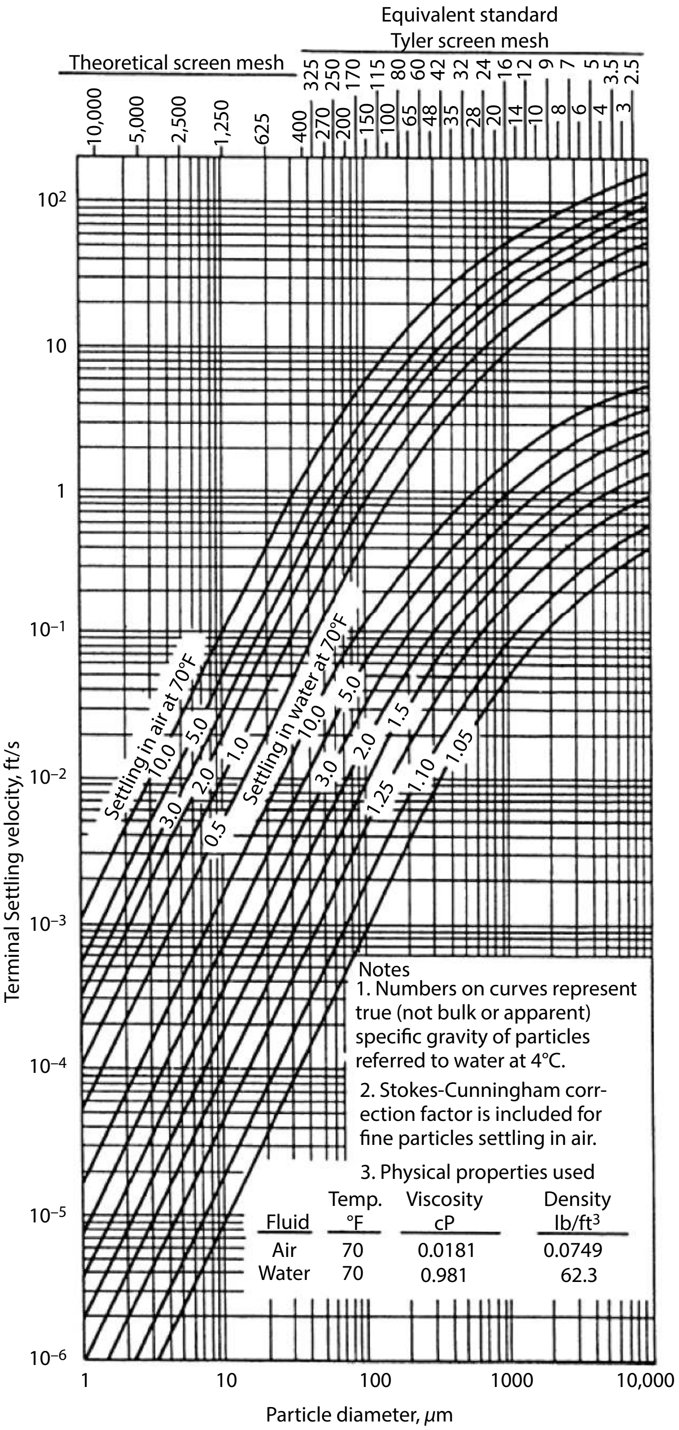


Figure 18.10 Terminal settling velocity of particles in water (by permission from Lapple, C. E., Fluid and Particle Mechanics, 1st ed. Univ. of Delaware. Newark, 1954).

particles falling through water. If a particle (liquid or solid) is falling under the influence of gravity through a vapor stream, the particle will continue to fall until, or unless the vapor flow rate is increased up to or beyond the terminal velocity value of the particle. If the vapor velocity exceeds this, then the particle will be carried along with the vapor (entrained).



Notes

1. Numbers on curves represent true (not bulk or apparent) specific gravity of particles referred to water at 4°C.
2. Stokes-Cunningham correction factor is included for fine particles settling in air.
3. Physical properties used

Fluid	Temp. °F	Viscosity cP	Density lb/ft³
Air	70	0.0181	0.0749
Water	70	0.981	62.3

Figure 18.11 Terminal velocities of spherical particles of different densities settling in air and water at 70°F under the action of gravity. To convert ft/s to m/s, multiply by 0.3048. From Lapple *et al.*, Fluid an Particle Mechanics, Univ. of Delaware, 1951, p. 292 (by permission from Perry, J. H., Chemical Engineers Handbook, 6th ed., McGraw-Hill Book, Co. Inc., pp. 5–67, 1984).

Pressure Drop

Pressure drop through gravity settlers is usually extremely low due to the very nature of the method of handling the problem.

Figure 18.11 is convenient for quick checks of terminal settling velocities of solid particles in air and in water [8].

18.6 American Petroleum Institute's Oil Field Separators

The American Petroleum Institute Manual on Disposal of Refinery Wastes provides specific design and construction standards for API-separators that are often used in oilfield waste disposal [14].

18.7 Liquid/Liquid, Liquid/Solid Gravity Separations, Decanters, and Sedimentation Equipment

Lamella Plate Clarifier: This angle plate gravity separator (Figure 18.12) removes suspensions of solids from a dilute liquid. The unit is more compact than a box-type settler due to the increased capacity achieved by the multiple parallel plates. The concept is fairly standard (U.S. Patent 1,458.805—year 1923) but there are variations in some details. For effective operation, the unit must receive the mixture with definite particles having a settling velocity. The units are not totally effective for flocculants or coagulated masses that may have a tendency to be buoyant.

Flow through the plates must be laminar, and the critical internal areas are [15–17] as follows:

- Distribution of the inlet flow.
- Flow between each parallel plate surface.
- Collection of clarified water.

Because these units are essentially open, the "standard" design does not fit into a closed process system without adding some enclosures, and certainly is not suitable for a pressurized condition.

Although the main suspended flow is through a top mixing chamber, the mixture flows around the angled (avg. 55°) parallel plate enclosure and begins its settling path from the bottom of the plates, flowing upward while depositing solids that slide countercurrent into the bottom outlet. The purified liquid flows overhead and out the top collecting trough.

Manufacturers should be contacted for size/rating information, because the efficiency of a design requires proper physical property information as well as system capacity and corrosion characteristics.

Thickeners and Settlers

Generally, large volume units for dewatering, settling of suspensions, and thickening and concentration of solids, and clarification must be designed by the specific manufacturer for the process conditions and physical properties. Some typical processes involving this class of equipment are lime slurrying, ore slurrying, ore clarifying, wastewater clarification and thickening. A good summary discussion is by Fitch [18].

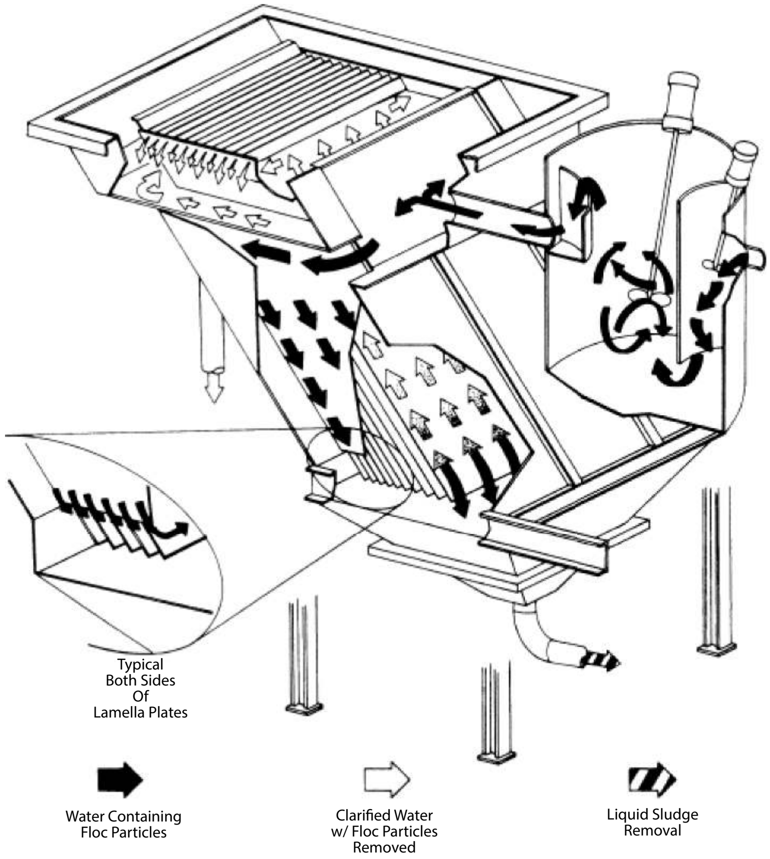


Figure 18.12 Lamella clarifier (by permission from Graver Water Div., of the Graver Co.).

18.8 Horizontal Gravity Settlers or Decanters, Liquid/Liquid

Many processes require the separation of immiscible liquid/liquid streams; that is, water/hydrocarbon. The settling unit must be of sufficient height (diameter) and length to prevent entrainment of the aqueous phase into the hydrocarbon and vice versa. Horizontal units are usually best for settling and possibly vented units for decantation (but not always).

Residence time of the mixture in the vessel is a function of the separation or settling rate of the heavier phase droplets through the lighter phase. Most systems work satisfactorily with a 30 min. to 1 h residence time, but this can be calculated [19]. After calculation, give a reasonable margin of extra capacity to allow for variations in process feed rate and in the mixture phase composition.

From Stokes' law, the terminal settling velocity:

$$V_t = \frac{gD_p^2(\rho_p - \rho)}{18\mu}, \text{ ft/s.} \tag{18.33}$$

μ = viscosity of surrounding fluid, lb/ft s
D_p = diameter of particle, ft

For assumed spherical particles in a surfactant-free system [9], the minimum particle diameter for many fine dispersions is 100 μm; however, Hooper and Jacobs [20] have reviewed a wide variety of liquid drop data and suggest that a good choice is 150 μm or 0.15 cm or 0.0005 ft. This is also the particle size used in the API Design Manual [14]. Using too large an assumed particle diameter will cause the settler unit to become unreasonably small.

Assuming a horizontal unit, as illustrated in Figure 18.13, has a segment of a circle equal to 25% to 75% [18, 19] of the circular area (the highest of this segment will be about 30% to 70% of the diameter), then height of the interface will be as follows [19].

$$H/D = \frac{0.8A}{\left(\pi D^2/4\right)} + 0.1, \text{ft} \tag{18.34}$$

or

$$h_c = \frac{38.4A}{(\pi D)} + 1.2D, \text{in.} \tag{18.35}$$

where

H = height of segment of a circle, ft
h_c = height of segment of a circle, in
D = diameter of vessel, ft
A = area of segment of circle, ft^2

The average volumetric residence time in the settler is:

$$t_{avg} = 7.48\ (V_{set}/F), \text{ min} \tag{18.36}$$

V_{set} = active volume of settler occupied by one of the phases, ft^3
F = flow rate of one phase, gpm
t_{avg} = average residence time based on liquid flow rate and vessel volume, min

The minimum residence time as determined by Stokes' law terminal settling velocity is given as follows:

$$t_{min} = h_c/v_t, \text{ min} \tag{18.37}$$

h_c = height of segment of circle, in.
v_t = terminal settling velocity of a particle under the action of gravity, in/min

Average residence time related to minimum residence time is:

$$t_{ave} = (f)\ (t_{min}) \tag{18.38}$$

f = factor relating average velocity to maximum velocity

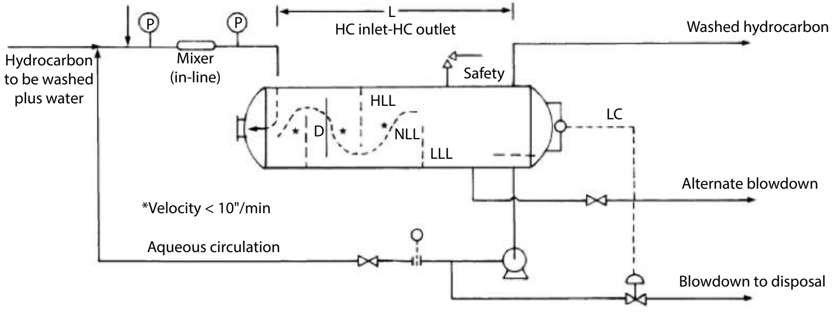


Figure 18.13 Settler vessel; runs full (adapted by permission from Abernathy, M.W., Encyclopedia of Chemical Processing and Design, Vol. 25, J. J. McKetta, ed., Marcel Dekker, 1987, p. 77 and Hydrocarbon Proc., Sept. 1977, p. 199 [15] and Ludwig [21].

This relationship is related to the viscosities of the hydrocarbon and aqueous phases at the interface. Based on data from different systems:

$$f = 2.0 \text{ (use for design)}$$

The active volume occupied by either phase is:

$$V = AL, \text{ ft}^3 \tag{18.39}$$

L = length of vessel, ft, inlet to outlet

$$h_c = 7.48\ ALv_t/(fF) \tag{18.40}$$

For an aqueous-hydrocarbon or organic solvent mixture, the top layer will be hydrocarbon, with the aqueous layer droplets settling through the hydrocarbon. The terminal velocity is:

$$v_{hc} = 12.86\ (\Delta SpGr)/\mu_{hc}, \text{ in./min} \tag{18.41}$$

where

v_{hc} = terminal settling velocity of aqueous droplets in hydrocarbon phase in top of vessel, in./min
$\Delta SpGr$ = differences in specific gravity of the particle and surrounding fluid
μ_{hc} = viscosity of surrounding fluid, cP

Height of hydrocarbon layer to the interface:

$$h_t = (7.48)\frac{(A_t L v_{hc})}{(f_{hc} F_{hc})} \tag{18.42}$$

where

F_{hc} = hydrocarbon phase flow, gpm

Also,

$$h_t = 38.4\ A_t/(\pi D) + 1.2D \tag{18.43}$$

where

h_t = height of continuous hydrocarbon phase in the top of vessel, in.

then,

$$A_t = \frac{1.2D}{\left[\dfrac{7.48 L v_{hc}}{(f_{hc} F_{hc})} - \dfrac{38.4}{(\pi D)}\right]} \tag{18.44}$$

where

A_t = cross-sectional area at top of vessel occupied by the continuous hydrocarbon phase, ft^2
A_b = cross-sectional area at bottom of vessel occupied by continuous aqueous phase, ft^2

For the bottom aqueous phase: hydrocarbon droplets settle out of the continuous aqueous phase. The terminal velocity is for hydrocarbon droplets:

$$v_{aq} = 12.86(\Delta SpGr)/\mu_{aq}, \text{ in./min} \tag{18.45}$$

v_{aq} = terminal settling velocity of hydrocarbon droplets in aqueous phase in bottom of vessel, in./min
μ_{aq} = viscosity of aqueous phase, cP

Height of Aqueous Layer to the Interface

$$h_b = (7.48)\ (A_b L v_{aq})/(f_{aq} F_{aq}) \tag{18.46}$$

Also,

$$h_b = 38.4 A_b / \pi D + 1.2 D \tag{18.47}$$

where
h_b = height of continuous aqueous phase in bottom of vessel, in.
A_b = cross-sectional area at bottom of vessel occupied by continuous aqueous phase, ft^2
f_{aq} = factor relating aqueous phase average velocity to maximum velocity
F_{aq} = aqueous phase flow rate, gpm
L = length of vessel from hydrocarbon inlet to hydrocarbon outlet, ft
v_{aq} = terminal settling velocity of hydrocarbon droplets in aqueous phase in bottom of vessel, in/min

$$A_b = \frac{1.2 D}{\left[\dfrac{7.48 L v_{aq}}{(f_{aq} F_{aq})} - \dfrac{38.4}{(\pi D)}\right]} \tag{18.48}$$

Optimum Vessel Diameter

Assume 20% cross-sectional area is occupied by an emulsion and is recognized as a "dead volume". This is actually the height over which the interface level will vary during normal operations [19].

$$A_t + A_b = 0.8\pi D^2/4 \tag{18.49}$$

where
A_t = cross sectional area at top of vessel occupied by continuous hydrocarbon phase, ft^2
A_b = cross sectional area at bottom of vessel occupied by continuous aqueous phase, ft^2
D = diameter of vessel, ft

$$D = \pm\left[\frac{a}{2} \pm \frac{\sqrt{(a^2 - 4b)}}{2}\right]^{0.5}, \text{ ft.} \tag{18.50}$$

where

$$a = (1.889)\,(v_{hc} f_{aq} F_{aq} + v_{aq} f_{hc} F_{hc}) \big/ (r v_{hc} v_{aq}) \tag{18.51}$$

$$b = (3.505)\,(f_{hc} F_{hc} f_{aq} F_{aq}) \big/ (r^2 v_{hc} v_{aq}) \tag{18.52}$$

where

r = economical vessel radius (= L/D), ft

v_{hc} = terminal settling velocity of aqueous droplets in hydrocarbon phase in top of vessel in/min.

18.9 Modified Method of Happel and Jordan

This method is a modification of the earlier method [23] by Abernathy [19], as follows, and can be less conservative than the original method [22, 23]. A basic assumption is that particles must rise/fall through one-half of the drum vertical cross-sectional area [19].

$$t = h/v$$

$$t = (1/2)(7.48)[0.8\pi D^2 L/4]F_t \tag{18.53}$$

where

D = diameter of vessel, ft.

L = length of vessel from hydrocarbon inlet to hydrocarbon outlet, ft.

F_t = flow rate of both phases, gpm

v_t = v = terminal settling velocity of a particle under the action of gravity, in./min

This assumes 20% of the cross-sectional even as "dead volume". The height from the interface can be determined by combining the above equations:

$$h = (0.748)\pi D^2 L v/F_t \tag{18.54}$$

The height for each interface is:

$$h_t = (0.748)\pi D^2 L h_{hc}/F_t \tag{18.55}$$

where

h_t = height of continuous hydrocarbon phase in the top of the vessel, in.

$$h_b = (0.748)\pi D^2 L v_{aq}/F_t \tag{18.56}$$

where

h_b = height of continuous aqueous phase in the bottom of the vessel, in

$$A_t = \left[\frac{(0.748)\pi D^2 L v_{hc}}{F_t} - 1.2D\right]\frac{\pi D}{38.4} \tag{18.57}$$

where

A_t = cross sectional area at top of vessel occupied by continuous hydrocarbon phase, ft^2.

$$A_b = \left[\frac{(0.748)\pi D^2 L v_{aq}}{F_t} - 1.2D\right]\frac{\pi D}{38.4} \tag{18.58}$$

where

A_b = cross sectional area at bottom of vessel occupied by continuous aqueous phase, ft^2.

Example 18.3 Horizontal Gravity Settlers

Using the data from Sigales [24] and following the design of [19], data for propane/caustic wash is as follows:

F_{hc} = 95 gpm
F_{aq} = 39 gpm
v_{aq} = 5 in./min
v_{hc} = 120 in./min
r = 3.4

The terminal (highest calculated) settling velocity of the aqueous droplet in/through the hydrocarbon phase is:

$$v_{hc} = (1.2)\ (5 \text{ in./min})\ (95 \text{ gpm}/39 \text{ gpm}) = 14.6 \text{ in./min}$$

Because this is more than the 10 in./min recommended earlier, then use:

$$v_{hc} = 10 \text{ in./min}$$

Assume for design: $f_{hc} = f_{ag} = 2$ (from earlier discussion).

Then, using Eq. 18.51

$$a = (1.889)\frac{[(10)(2)(39)+(5)(2)(95)]}{[(3.4)(10)(5)]} = 19.22$$

From Eq. 18.52,

$$b = \frac{(3.505)(2)(95)(2)(39)}{[(3.4)^2(10)(5)]} = 89.87$$

From Eq. 18.50,

$$D = \pm\left[\frac{19.22}{2} \pm \frac{\sqrt{(19.22^2 - 4(89.87))}}{2}\right]^{0.5}, \text{ft.}$$

D = 3.34 ft or −2.83 ft (latter is an unreal negative number, so use 3.34 ft)
A_t = Area of segment at top of vessel

Substituting into Eq. 18.44 gives:

$$A_t = \frac{1.2 \times 3.34}{\left[\frac{7.48 \times 3.4 \times 3.34 \times 10}{(2 \times 95)} - \frac{38.4}{(\pi \times 3.34)}\right]} = 4.945\,\text{ft}^2$$

Using: L/D = 3.4 for the bottom segment of the vessel, aqueous layer

$$A_b = \frac{1.2 \times 3.34}{\left[\frac{7.48 \times 3.4 \times 3.34 \times 5}{(2 \times 39)} - \frac{38.4}{(\pi \times 3.34)}\right]} = 2.2448\,\text{ft}^2$$

Then, using Eq. 18.42

$$h_t = \frac{7.48(4.945)(3.4 \times 3.34)(10)}{(2 \times 95)} = 22.1\,\text{in.}$$

$$h_b = \frac{7.48\,A_b\,L\,v_{aq}}{(f_{aq}\,F_{aq})}$$

$$h_b = \frac{7.48(2.2488)(3.4 \times 3.34)(5)}{(2 \times 39)} = 12.2\,\text{in.}$$

Then,

$$\frac{h_t}{D} = \frac{(22.1)}{(12)(3.34)} \times 100 = 55\%$$

$$\frac{h_b}{D} = \frac{(12.2)}{(12)(3.34)} \times 100 = 30\%$$

Since h_t and h_b are between 30% and 70% of the diameter, the solution is acceptable.

In summary:

Design calculation	**Practical design use**
Diameter 3.34 ft (40.08 in.)	3.5 ft (42 in.) or 3.83 ft (46 in.)
Length HC inlet/outlet: 11 ft	12 or 14 ft

Abernathy [19] has compared several design methods as follows:

	Sigales	**This method**	**Modified Happel**	**Happel**	**Rule of thumb**
Diameter	2.67 ft	3.34 ft	3.36 ft	4.01 ft	4.1 ft
h_t	10 in.	22 in.	22.6 in.	24 in.	32.5 in.
h_b	8 in.	12 in.	11.3 in.	24 in.	16.7 in.
Interface	14 in.	6 in.	6.4 in.	0 in.	0 in.
HC residence time	1.1 min	4.4 min	4.6 min	6.8 min	10 min

18.10 Decanter

In most general applications, a decanter is a continuous gravity separation vessel that does not run full, as contrasted to a settler that usually runs full, with one stream exiting at or near the top of a horizontal vessel [25]. For most decanters, one phase of a two-phase mixture overflows out of the vessel (Figure 18.14). The concept of the decanter involves the balancing of liquid heights due to differences in density of the two phases, as well as settling velocity of the heavier phase falling through the lighter, or the lighter rising through the heavier.

Terminal settling velocity [25]:

$$v_d = gd^2 \frac{(\rho_d - \rho_c)}{18\mu_c}, \text{ ft/s} \tag{18.59}$$

where

v_d = terminal settling velocity or a droplet, ft/s
g = acceleration due to gravity, 32.2 ft/s^2
d = droplet diameter, ft(1 ft = 304, 800 μm, or 1 μm = 0.001mm)
ρ_d = density of fluid in the droplet, lb/ft^3
ρ_c = density of fluid continuous phase, lb/ft^3
μ_c = viscosity of the continuous phase, lb/ft s

Note: 1 cP = 6.72×10^{-4} lb/ft s
1 μm = micron = 10^{-6} m = 39.37×10^{-6} in.

For a decanter that operates under gravity flow with no instrumentation flow control, the height of the heavy phase liquid leg above the interface is balanced against the height of one light phase above the interface [8]. Figures 18.14 and 18.15 illustrate the density relationships and the key mechanical details of one style of decanter.

The same results can be achieved with internal flat plate baffles and outlet nozzles.

$$(z_h - z_i)\rho_h = (z_l - z_i)\rho_L \tag{18.60}$$

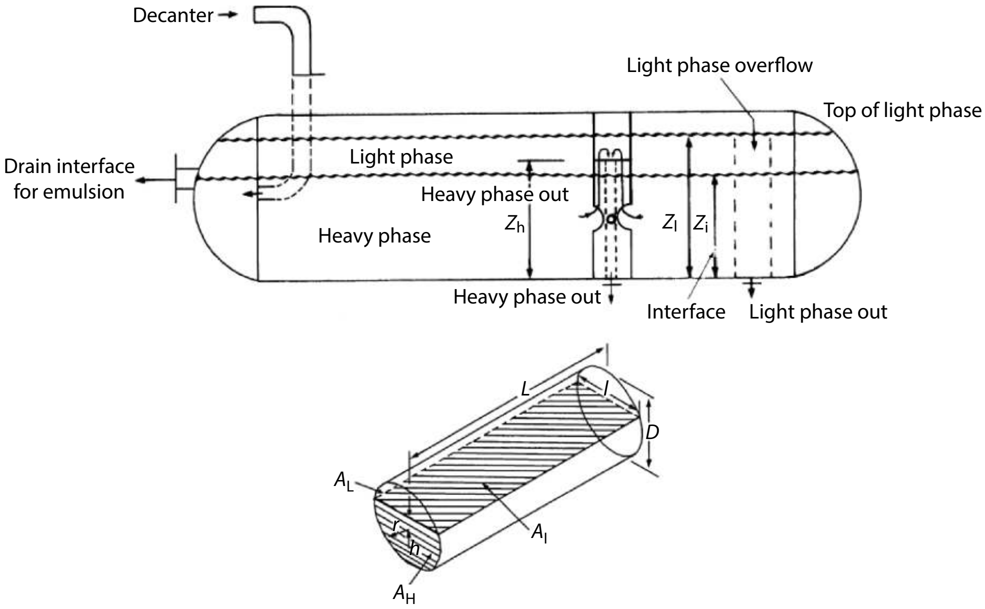


Figure 18.14 Gravity decanter basic dimensions (adapted by permission from Schweitzer, P.A., Handbook of Separation Techniques for Chemical Engineers, McGraw-Hill Book Co., 1979 [25]).

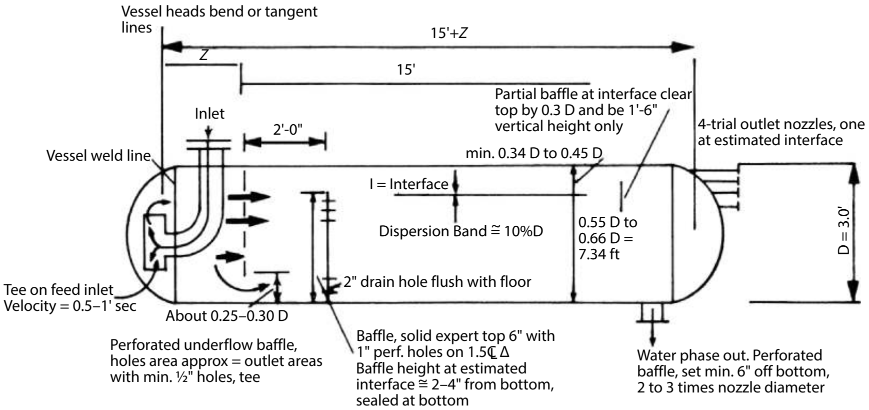


Figure 18.15 Decanter.

where

z_h = heavy phase outlet dimension from bottom of horizontal decanter
z_i = interface measured from bottom
z_l = light phase outlet measured from bottom of decanter

Droplet diameter, when other data is not available:

$$d = 150\ \mu\text{m}\ (0.0005\ \text{ft})$$

Schweitzer [25] recognizes that this is generally on the safe side, because droplets generated by agitation range 500 to 5000 μm, turbulent droplet range 200 to 10,000 μm. Due to limitations of design methods, decanters sized for droplets larger than 300 μm often result in being too small to work properly.

The continuous phase moves through the vessel on a uniform flow equal to the overflow rate. To identify which is the continuous phase (from Selker and Sleicher by Schweitzer),

$$\theta = \frac{Q_L}{Q_H}\left(\frac{\rho_L \mu_H}{\rho_H \mu_L}\right)^{0.3} \tag{18.61}$$

θ	*Result*
<0.3	light phase always dispersed
0.3–0.5	light phase probably dispersed
0.5–2.0	phase inversion probable, design for worst case
2.0–3.3	heavy phase probably dispersed
>3.3	heavy phase always dispersed

where

Q_L = volumetric flow rate, ft^3/s, light phase
Q_H = volumetric flow rate, ft^3/s, heavy phase
ρ_L = density of light phase fluid, lb/ft^3
ρ_H = density of heavy phase fluid, lb/ft^3
μ_H = viscosity of heavy phase, lb/ft s
μ_L = viscosity of light phase, lb/ft s

There is a dispersion band through which the phases must separate. Good practice [25] normally keeps the vertical height of the dispersed phase, $H_D < 10\%$ of decanter height (normally a horizontal vessel), and

$$\frac{1}{2} H_D \frac{A_I}{Q_D} > 2-5\,\text{min}.$$

where

A_I = area of interface assuming flat interface, ft^2
H_D = height of the dispersion band, ft
Q_D = volumetric flow, dispersed phase, ft^3/s
D = decanter diameter, ft

Horizontal vessels as cylinders are generally more suitable for diameters up to about 8 ft than other shapes, or vertical, due in part to the increased interfacial area for interface formation. For a horizontal drum (see Figure 18.14):

$$I = 2(r^2 - h^2)^{1/2} \tag{18.62}$$

$$A_I = IL \tag{18.63}$$

where

h = distance from center to given chord of a vessel, ft
I = width of interface, ft
L = decanter length, ft
r = vessel radius, ft

$$A_L = 1/2\,\pi r^2 - h(r^2 - h^2)^{1/2} - r^2 \text{ arc sin}(h/r) \tag{18.64}$$

Radians = (degrees) (π/180)

$$A_H = \pi r^2 - A_L \tag{18.65}$$

$$D_L = \frac{4A_L}{(I+P)} \tag{18.66}$$

where D_L = diameter for light phase, ft.

$$D_L = \frac{4A_H}{(I+2\pi r-P)} \tag{18.67}$$

where

A_L = cross-sectional area allotted to light phase, ft^2
A_H = cross-sectional area allotted to heavy phase, ft^2
P = 2r arc cos(h/r)

Degree of turbulence [25]:

$$N_{Re} = \frac{v_c D_H \rho_c}{\mu_c} \tag{18.68}$$

where

c = continuous phase

D_H = hydraulic diameter, ft = 4 (flow area for the phase in question/wetted perimeter of the flow channel)

v_c = velocity down the flow channel for continuous phase, ft/s

Guidelines for Successful Decanters [25]:

Re	Results
<5000	Little problem
5000–20,000	Some hindrance
20,000–50,000	Major problem may exist
Above 50,000	Expect poor separation

Velocities of both phases should be about the same through the unit. By adjusting mechanical internals, a ratio of <2:1 is suggested (internals do not need to be equal) [25]. Velocities for entrance and exit at the vessel nozzle should be low, in the range of 0.5 to 1.5 ft/s. The feed must not "jet" into the vessel, and should be baffled to prevent impingement in the main liquid body, keeping turbulence to an absolute minimum to none. Baffles can/should be placed in the front half of the unit to provide slow flow of the fluids either across the unit or up/down paths followed by the larger stilling chamber, before fluid exits (see Figure 18.15).

Example 18.4

Decanter, using the method of Schweitzer

A plant process needs a decanter to separate oil from water [25]. The conditions are:

Oil flow = 8500 lb/h
ρ = 56 lb/ft^3
μ = 9.5 cP

Water flow = 42,000 lb/h
ρ = 62.3 lb/ft^3
μ = 0.71 cP

Units conversion:

Q_{oil} = (8500/(56) (3600) = 0.0421 ft^3/s
μ_{oil} = (9.5) (6.72 × 10^{-4}) = 63.84 × 10^{-4} 1b/ft-s
Q_{water} = 42,000/(62.3) (3600) = 0.1873 ft^3/s
μ_w = (0.71) (6.72 × 10^{-4}) = 4.77 × 10^{-4} 1b/ft-s

Checking dispersed phase, Eq. 18.61:

$$\theta = \frac{0.0421}{0.1873}\left[\frac{(56)(4.77\times10^{-4})}{(62.3)(63.84\times10^{-4})}\right]^{0.3}$$
$$= 0.09997$$

Therefore, light phase is always dispersed since θ is less than 0.3.

Settling rate for droplets of oil through water from Eq. 18.20

Assume droplet size is d = 0.0005 ft (150 μm), as earlier discussed and using Eq. 18.20.

$$V_{oil} = \frac{32.2 \times 0.0005^2 \times (56 - 62.3)}{(18 \times 4.77 \times 10^{-4})}$$
$$= -0.006 \text{ ft/s.}$$

The (–) sign means the oil rises instead of settles.

Overflow rate:

Assume I (Figure 18.14) is 80% diameter, D, of vessel and that L/D = 5.

where
L = length of the vessel, ft.

Then,

$$Q_c/A_I < v_d$$
$$A_I = IL = (0.8D)\ (5D) = 4D^2\text{, then,}$$
$$Q_c/4D^2 < v_d$$

and:

$$D \geq 1/2\ (Q_c/v_d)^{1/2} \geq 1/2\ (0.187/0.006)^{1/2}$$
$$D = 2.79 \text{ ft}$$

where
v_d = terminal settling velocity of a droplet, ft/s
Length, L = 5D = 5 (2.79) = 14ft

Interface Level

Assume that the interface is being held one foot below top of vessel to prevent interface from reaching the top oil outlet.

Then,
h = 0.5 ft
r = 3.0/2 = 1.5 ft

From Eq. 18.62

$$I = 2(r^2 - h^2)^{1/2} = 2[(1.5)^2 - (0.5)^2]^{1/2} = 2.828 \text{ ft}$$

And from Eq. 18.64

$$\begin{aligned} A_{oil} &= (1/2)\ (\pi)\ (1.5)^2 - 0.5[(1.5)^2 - (0.5)^2]^{1/2} - (1.5)^2 \text{ arc sin. } (0.5/1.5) \\ &= 3.534 - 0.707 - 0.765 \\ &= 2.062 \text{ ft}^2 \end{aligned}$$

Note: In radians: Arc sin (0.5/1.5) = (19.47/180)π = 0.3398

$$A_{water} = \pi(1.5)^2 - A_{oil} = \pi(2.25) - 2.062 = 5.01 \text{ ft}^2$$
$$P = 2(1.5)\ [\text{arc cos } (0.5/1.5)] = 3.69 \text{ ft}^2$$

$$\text{Area interface, } A_I = (I)\,(L) = (2.828)\,(14) = 39.59\ \text{ft}^2$$

Secondary settling:

Continuous phase water droplets to resist the oil overflow rate if it gets on the wrong side of interface.

$$v_{water} \le Q_{oil}/A_I = 0.0421/39.59 = 0.00106\ \text{ft/s}$$

Then, from settling-velocity equation, the particle diameter, d is:

$$d = \left[\frac{18 v_t \mu}{g_c(\rho - \rho_s)}\right]^{0.5}$$

$$d = \left[\frac{18 \times 6.384 \times 10^{-4} \times 0.00106}{32.17 \times (62.3 - 56)}\right]^{0.5}$$

$$d = 0.0007752\ \text{ft}, (236\ \mu\text{m})$$

Note: 1 μm = 39.37×10^{-6} in. = 3.2808×10^{-6} ft.

Checking coalescence time: Assume H_D = height of dispersion band = 10% of D = 0.3 ft.

Time available to cross the dispersed band = $1/2(H_D A_I/Q_D)$ should be >2 to 5 min

= 1/2[(0.279) (39.59)/(0.0421)]
= 131 s, which is 2.2 min

Should be acceptable, but is somewhat low.

Then

$$D_{oil} = 4(2.062)/(2.828 + 3.69) = 1.265\ \text{ft}$$

$$V_{oil} = Q_{oil}/A_{oil}$$

$$V_{oil} = 0.0421/(2.062) = 0.0204\ \text{ft/s}$$

The Reynolds number is:

$$N_{Re} = \frac{D_{oil} V_{oil} \rho_{oil}}{\mu_{oil}}$$

$$N_{Re,\,oil} = \frac{1.2654 \times 0.0204 \times 56}{63.84 \times 10^{-4}} = 226.4$$

$$D_{water} = 4(5.01)/[2.828 + 2\pi(1.5) - 3.69] = 2.34\ \text{ft}$$

$$v_{water} = \frac{Q_{water}}{A_{water}},\ \text{ft/s}$$

$$= 0.187/5.01 = 0.0373\ \text{ft/s}$$

The Reynolds number is:

$$N_{Re(water)} = \frac{(0.0373)(2.34)(62.3)}{4.77 \times 10^{-4}} = 11{,}399$$

d = droplet diameter, ft.

The degree of turbulence would be classified as acceptable, but the unit must not be increased in capacity for fear of creating more water phase turbulence.

18.11 Impingement Separators

As the descriptive name suggests, the impingement separator allows the particles to be removed to strike some type of surface. This action is better accomplished in pressure systems where pressure drop can be taken as a result of the turbulence which necessarily accompanies the removal action.

Particle removal in streamline flow is less efficient than for turbulent flow, and may not be effective if the path of travel is not well baffled.

The "target" efficiency for impingement units expresses the fraction of the particles in the entraining fluid, moving past an object in the fluid, which impinge on the object. The target efficiencies for cylinders, spheres, and ribbon-like particles are given for conditions of Stokes' Law in an infinite fluid by Figure 18.16.

If the particles are close enough together in the fluid to affect the path of each other, then Figure 18.16 gives conservative efficiencies. For particles differing considerably from those given in the curves, actual test data should be obtained.

There are basically three construction types for impingement separators:

1. Wire mesh
2. Plates (curved, flat or special shaped)
3. Packed Impingement Beds

Knitted Wire Mesh

A stationary separator element of knitted small diameter wire or plastic material is formed of wire 0.003 in. to 0.016 in. (or larger) diameter into a pad of 4, 6, or 12 in. thickness and serves as the impingement surface for liquid particle separation. Solid particles can be separated, but they must be flushed from the mesh to prevent plugging. Although several trade name units are available, they basically perform on the same principle, and have very close physical characteristics. Carpenter [26] presented basic performance data for mesh units. Figure 18.17 shows a typical eliminator pad.

Figure 18.18 pictorially depicts the action of the wire mesh when placed in a vertical vessel. Referring to Figure 18.18, the typical situation represents a vapor disengaging from a liquid by bursting bubbles and creating a spray of liquid

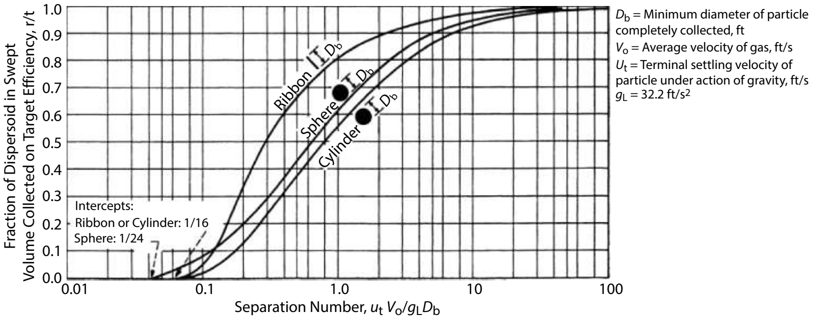
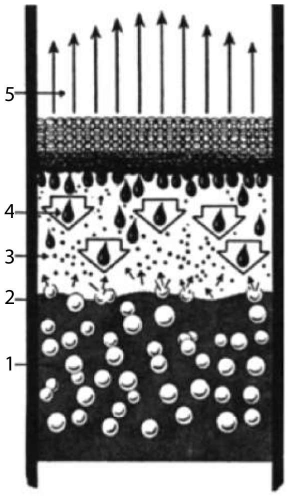


Figure 18.16 Target efficiencies for spheres, cylinders, and ribbon-like particles (by permission from Perry, J. H., Chemical Engineer's Handbook, 3rd ed., McGraw-Hill Company, 1950 [5]).

Figure 18.17 Eliminator pad (courtesy of Otto H. York Co.).



When a gas is generated in, or passes through, a liquid (1), the gas on bursting from the liquid surface (2) carries with it a fine spray of droplets–liquid entrainment–which are carried upward in the rising gas stream (3). As the gas passes through the mist eliminator, these droplets impinge on the extensive surface of the wire, where they are retained until they coalesce into large drops. When these liquid drops reach sufficient size, they break away from the wire mesh (4) and fall back against the rising gas stream. In this way, the entrained droplets are literally "wiped out" of the gas which, freed from liquid entrainment, (5) passes on unhindered through the mesh.

Figure 18.18 Diagram of action of wire mesh in liquid–vapor separation (courtesy of Metal Textile Corp. Bulletin ME 9-58).

particles of various sizes. Many of these particles are entrained in the moving vapor stream. The largest and heaviest particles will settle by gravity downward through the stream and back to the bottom of the vessel or to the liquid surface. The smaller particles move upward, and if not removed will carry along in the process stream. With wire mesh in the moving stream, the small particles will impinge on the wire surfaces; coalesce into fluid films and then droplets, run to a low point in their local system; and fall downward through the up-flowing gas stream when sufficiently large. The gas leaving is essentially free from entrained liquid unless the unit reaches a flooding condition.

For special applications the design of a mist eliminator unit may actually be an assembly in one casing of wire mesh and fiber packs/pads or in combination with Chevron style mist elements (see Figures 18.19a–c). This can result in greater recovery efficiencies for small particles and for higher flow rates through the combined unit. Refer to the manufacturers for application of these designs.

Mesh Patterns

There are several types of mesh available, and these are identified by mesh thickness, density, wire diameter and weave pattern. Table 18.9 identifies most of the commercial material now available. The knitted pads are available in any material that can be formed into the necessary weaves; this includes: stainless steels, monel, nickel, copper, aluminum, carbon steel, tantalum, Hastelloy, Saran, polyethylene, fluoropolymer, and glass multi-filament.

Capacity Determination

The usual practice in selecting a particular mesh for a given service is to determine the maximum allowable velocity and from this select a vessel diameter. In the case of existing vessels where mesh is to be installed, the reverse procedure is used, that is, determine the velocity conditions which will prevail and select a mesh to fit as close to the conditions as possible. The procedure is outlined below:

Allowable vapor velocity (mesh in horizontal position)

$$V_a = k\sqrt{\frac{\rho_L - \rho_v}{\rho_v}}$$

where

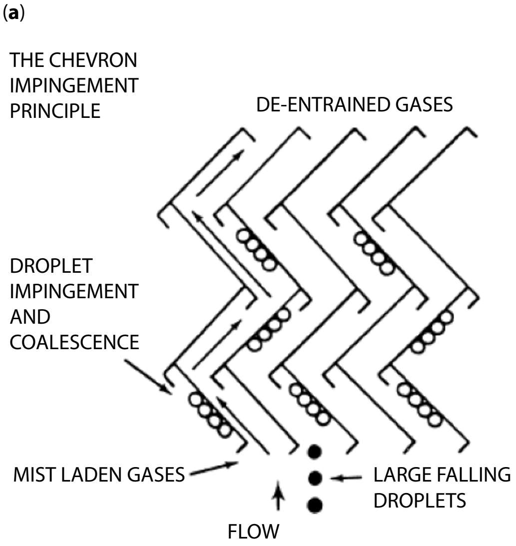


Figure 18.19a Separation/impingement action of Chevron-style mist eliminators. Flow is up the V-shaped plates assembly (courtesy of Bulletin KME-12, Koch Engineering Co.).

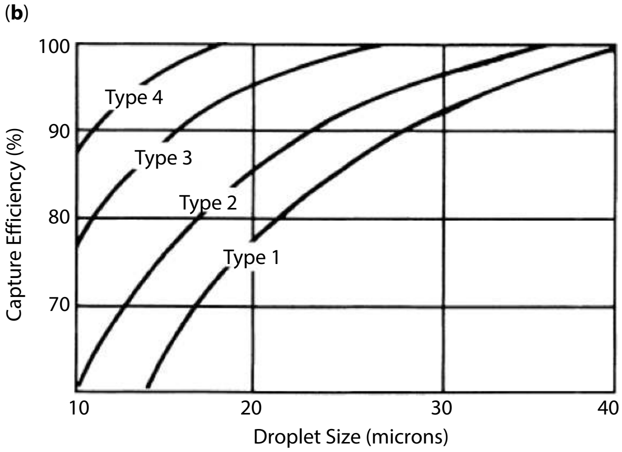


Figure 18.19b Capture efficiency vs. particle size for four standard York–Vane mist eliminators (by permission from Otto H. York, Co. Inc.).

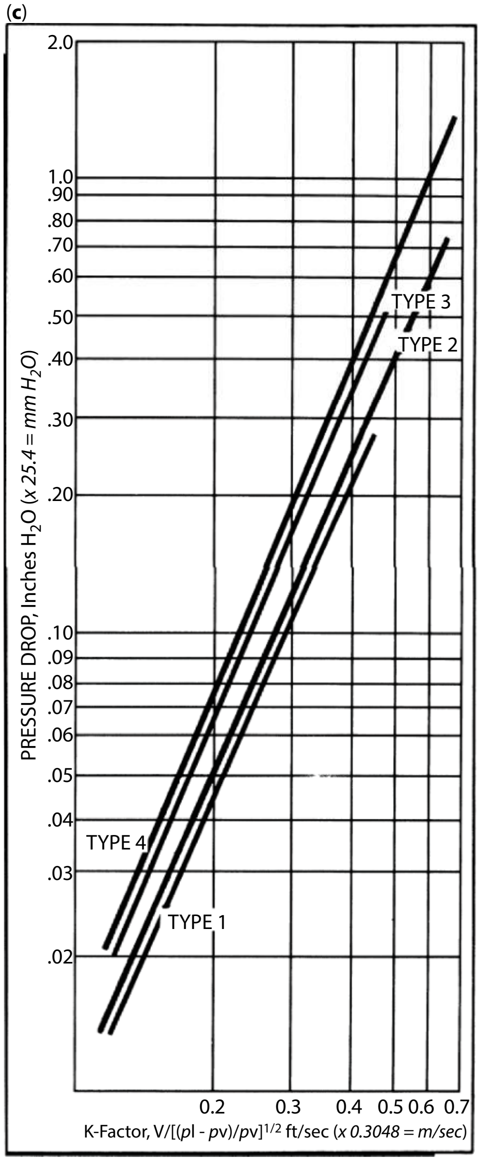


Figure 18.19c Pressure drop vs. K-factor for standard York–Vane mist eliminators, air–water system (by permission from Otto H. York, Co. Inc.).

Table 18.9 Identification of wire mesh types.

General type	Density (lb/ft^3)*	Surface area (ft^2/ft^3)	Thickness, (in.)**	Min. efficiency (wt. %)	Application
High Efficiency	12	115	4+	99.9+	Relatively clean, moderate velocity
Standard Efficiency	9	85	4+	99.5+	General purpose
Optimum Eff. Or VH Efficiency, and Wound type	13–14	120	4+	99.9+	For very high efficiency
Herringbone, High through-put or Low Density	5–7	65±	4–6+	99.0+	For services containing solids, or "dirty" materials

* If the mesh is made of nickel, monel, or copper, multiply the density values by 1.13, referenced to stainless steel.
** 4" is minimum recommended thickness; 6" is very popular thickness; 10" and 12" recommended for special applications such as fine mists, oil vapor mist.
Source: Compiled from [27, 28].

V_a = maximum allowable superficial vapor velocity across inlet face of mesh, ft/s
k = constant based on application, Table 18.10, average for free flowing system = 0.35 for 9–12 lb/ft^3 mesh
ρ_L = liquid density, lb/ft^3
ρ_v = vapor density lb/ft^3

For other mesh densities, use K_s [29] of 0.4 for 5 lb/ ft^3 mesh (high capacity), and 0.3 for plastic mesh such as Teflon® and polypropylene.

Table 18.10 "k" values of knitted mesh.

Bottom of mesh at least 12 in. above liquid surface		
Service conditions	**"k"**	**General type mesh**
Clean fluids, moderate liquid load, fits 90% of process situations	0.35 to 0.36	Standard
	0.35	High Efficiency
	0.25	Very High Efficiency
High viscosity, dirty suspended solids	0.40	Low density or Herringbone, high Through-put
Vacuum operations:		
2" Hg. Abs.	0.20	Standard or High Efficiency Plastic coated wire, or plastic strand
16" Hg. Abs.	0.27	
Corrosive Chemical	0.21	

Source: Compiled from various manufacturer's published data.
Note: k values for estimating purposes, not final design unless verified by manufacturer. Unless stated, all values based on stainless steel wire.

Fleximesh Design Manual [29] suggests "dry" mesh pressure drop of:

$$\Delta p_D = \left(\frac{f_c \, l a \rho_v \, V_s}{g_c \, \varepsilon^3} \right) \left(\frac{27.7}{144} \right) \tag{18.69}$$

$$\Delta p_T = \Delta p_D + \Delta p_L \tag{18.70}$$

For Δp_L see manufacturer's curves.

A rough approximation of operating mesh pressure drop is 1 in. H_2O or less. The calculated pressure drop at the maximum allowable velocity is close to 1.5 in. of H_2O. Therefore:

$$\Delta p_T = 1.5 \left(\frac{V_{act}}{V_{max}} \right)^2 \tag{18.71}$$

where

a = specific surface area, ft^2/ft^3
f_c = friction factor, dimensionless
g_c = gravitational constant, 32.174 $\left(\frac{lb_m}{lb_f} \right) \left(\frac{ft}{s^2} \right)$
l = wire mesh thickness, ft
Δp_D = pressure drop, no entrainment, in. H_2O
Δp_L = pressure drop, due to liquid load, in. H_2O
Δp_T = pressure drop, total across wet pad, in. H_2O
V_s = superficial gas velocity, ft/s
ε = void fraction of wire mesh, dimensionless
ρ_L = liquid density, lb/ft^3
ρ_v = vapor density, lb/ft^3
f = generally ranges 0.2 to 2 for dry mesh

Subscript:
act = actual
max = maximum

The correlation factor, k, is a function of the liquid drop size, liquid viscosity, liquid load, disengaging space, type of mesh weave, and so on. And k varies somewhat with system pressure; as pressure increases the k value decreases. The manufacturers should be consulted for final design k values for a system, because the wire style, size, and material also affect the value. For pressures below 30 psig, k = 0.35 avg., then above 30 psig, k value decreases with pressure with an approximate value of 0.30 at 250 psig and 0.275 at 800 psig. Certain values have been found satisfactory for estimating systems described in Tables 18.10 and 18.11.

For conditions of high liquid loading, use caution in design. Use the high velocities for very fine mist to remove the small particles, and use two mesh pads in series with the second mesh operating at a lower velocity to remove the larger drops re-entrained from the first mesh. Systems involving high viscosity fluids should be checked with the various manufacturers for their case history experience. Lower k values are used for systems with high vacuum, high viscosity liquids, low surface tension liquids and systems with very bad fouling conditions. Table 18.11 indicates the effect of disengaging height on the allowable k value. Similar relations should hold for other mesh densities.

Velocity Limitations. Very low velocities will allow particles to drift through the mesh and be carried out with the leaving vapor. Also, very high velocities will carry liquid to the top of the mesh, establish a "flooding" condition, and then re-entrain the liquid from the surface of the mesh. For most situations very good performance can be expected

Table 18.11 Variation of k with disengaging height.*

Disengaging height above mesh (in.)	Allowable k value
3	0.12
4	0.15
5	0.19
6	0.22
7	0.25
8	0.29
9	0.32
10	0.35
11	0.38
12	0.40
13	0.42
14	0.43

Source: By permission, from York [28].
Note: Values based on 12 lb/ft^3 wire mesh. Design practice normally does not exceed k of 0.4 even for higher disengaging height.

for all velocities from 30 to 100% of the optimum allowable design velocity. The minimum allowable safe design velocity is 10% of the value calculated by the equation. The flooding velocity of the mesh is usually about 120–140% of the maximum allowable velocity.

Generally, the maximum allowable velocities are lower under conditions of pressure, and higher under conditions of vacuum. The limits and ranges of each area being determined by the relative operating densities of the vapor and liquid, the nature of the entrainment, and the degree of separation required.

When the mesh is installed with the pad vertical or inclined, the maximum allowable velocity is generally used at 0.67 times the allowable value for the horizontal position.

Design Velocity. To allow for surges, variations in liquid load and peculiarities in liquid particle size and physical properties use:

$$V_D = 0.75\ V_a \tag{18.72}$$

for the design of new separators.

where

V_D = design vapor velocity, ft/s
V_a = maximum allowable superficial vapor velocity across inlet face of mesh, ft/s

When checking existing vessels to accept wire mesh, some variation may have to be accepted to accommodate the fixed diameter condition, but this is no great problem since the range of good operation is so broad.

The Design Parameter, K_s. The design parameter, K_s, is the Souders–Brown equation is an empirical parameter and is a key factor for the sizing the gas–liquid separators vessel diameter as well as for determination of the mist extractor diameter. Its value depends on several factors including

- Pressure
- Fluid properties (note temperature has a large impact on fluid properties).
- Separator geometry
- Vessel length and the liquid level (for horizontal separators).
- Steadiness of flow
- Inlet device design and performance
- Relative amounts of gas and liquid
- Most importantly—mist extractor type and design (e.g., mesh pad, vane pack, multi cyclone)

Figure 18.20 provides the relationship of K_s values for various droplet sizes and separator operating pressures for the gas gravity section. Typically, a K_s value is selected that will achieve removal of all entrained droplets larger than a specified target droplet diameter in the original design of the separator. For conventional separators the target droplet diameter is typically 150 μm, and for scrubbers the target droplet size should not exceed ~500 μm. This correlation can also be used to determine the performance of the gas gravity section based upon current operating conditions. The separator K_s value can be estimated from the actual velocity and fluid conditions, and the droplet size removed in the gas gravity section can be estimated from Figure 18.20.

Figure 18.20 K_s as a function of pressure and liquid droplet size for vertical separators with no mist extractor devices [89].

American Petroleum Institute (API 12J). The API 12 [81] recommends ranges of K_s-values for vertical and horizontal gas–liquid separators. These values are presented below:

Type	Height or length, ft (m)	Typical K_s range, ft/s	Typical K_s range, m/s
Vertical	5 (1.52)	0.12–0.24	0.037–0.073
	10 (3.05)	0.18–0.35	0.055–0.107
Horizontal	10 (3.05)	0.40–0.50	0.122–0.152
	Other Lengths	0.40–0.50 $(L/10)^{0.56}$	0.122–0.152 $(L/3.05)^{0.56}$

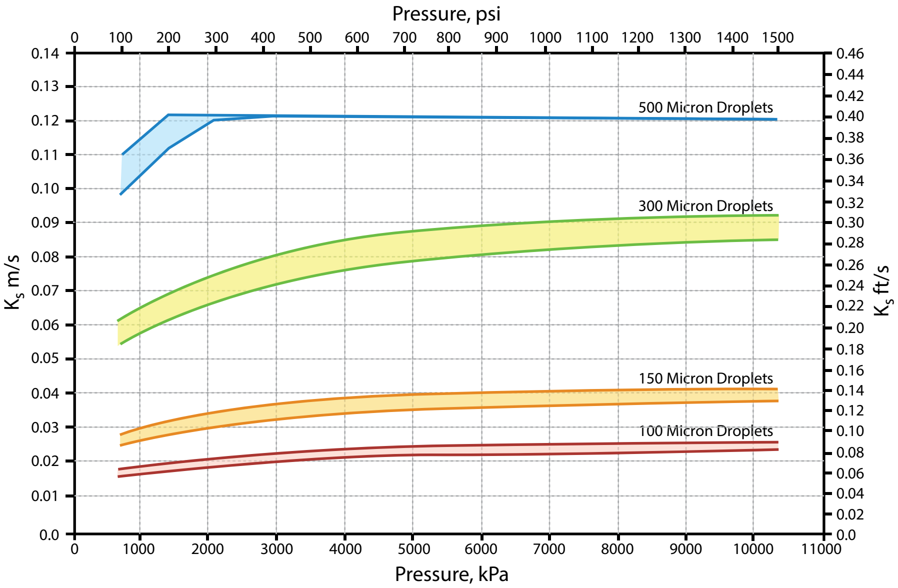


Figure 18.20 K_s as a function of pressure and liquid droplet size for vertical separators with no mist extractor devices [83].

Per API 12J "the maximum allowable superficial velocity, calculated from the above factors, is for separators normally having a wire mesh mist extractor. This rate should allow all liquid droplets larger than 10 μm to settle out of the gas. The maximum allowable superficial velocity or other design criteria should be considered for other type mist extractor. Mist extractor manufacturer's recommended minimum distances upstream and downstream of the wire mesh between gas inlet and outlet nozzles should be provided for full utilization of the mist extractor. These values assume separators are equipped with standard mesh pad mist extractors".

Efficiency. For most applications, the efficiency will be 98–99%+ as long as the range of operating velocity is observed. The typical performance curves for this type of material are given in Figures 18.19b, 18.21, and 18.22. For hydrocarbon liquid–natural gas system, guarantees are made that not more than 0.1 gal. of liquid will remain in the gas stream per million cubic feet of gas. Special designs using a 3-ft thick pad reduce radioactive entrainment to one part per billion [28].

For the average liquid process entrainment, the mesh will remove particles down to 4 - 6 μm at 95%+ recovery efficiencies (see Figure 18.21). Particles smaller than this usually require two mesh pads or the fiber pack style discussed later. Carpenter [26, 30] shows the calculated effect of decreasing particle size on percent entrainment removed at various linear velocities. For water particles in air at atmospheric pressure, the 8 μm particles are 99% removed at 3.5 ft/s, the 7 μm at 5 ft/s, and the 6 μm at 6.8 ft/s. An excellent performance may be obtained in most systems for velocities of 30–110% of calculated values [31].

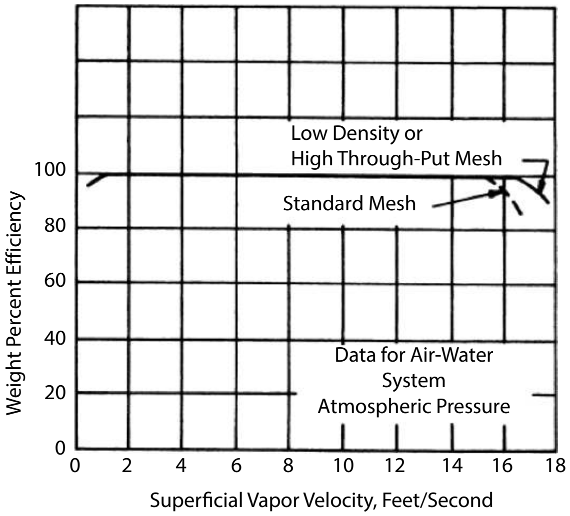


Figure 18.21 Typical wire mesh efficiency.

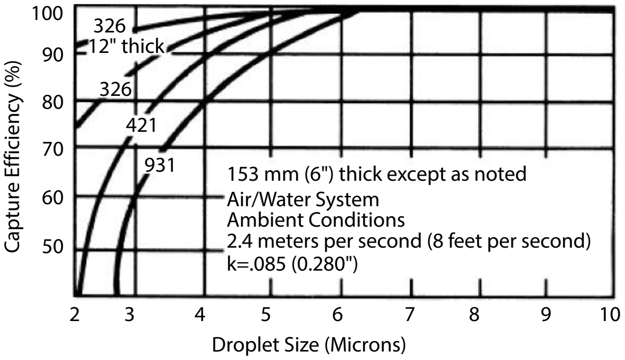


Figure 18.22 Capture efficiency vs. particle size for four types of DEMISTER® knitted mesh mist eliminators (by permission from Otto H. York Co. Inc.).

Pressure Drop. Pressure drop through wire mesh units is usually very low, in the order of 1-in. H_2O gauge for a 4-in. or 6-in. thick pad. For most pressure applications this is negligible. If solids are present in the particle stream, then solids build-up can become appreciable, and is usually the guide or indicator for cleaning of the mesh. A 12-in. pad may require a 3-in. H_2O drop. Figures 18.23 and 18.24 present the range of expected pressure drops for a spread of 3 to 1600 lb/h-ft^2 for liquid rates. Although this is for air-water system at atmospheric pressure it will not vary much unless the physical properties of the vapor and liquid deviate appreciably from this system, in which case the general Fanning equation can be used to approximate the pressure drop under the new conditions. Approximate values based upon air-water tests suggest these relations [27].

For the standard weave, 4 in. thick:

$$\Delta p = 0.2\, V_D^2 \rho_v, \text{ in. } H_2O \tag{18.73}$$

For the low-density weave (high through-put), 6 in. thick:

$$\Delta p = 0.12\, V_D^2 \rho_v \tag{18.74}$$

Installation. The knitted mesh separator unit may be placed in a pipe in which case a round flat rolled unit is usually used, or it may be placed in a conventional vessel. Although the vessel may be horizontal or vertical, the mesh must always be in a horizontal plane for best drainage. Some units in special situations have been placed at an angle to the horizontal, but these usually accumulate liquid in the lower portion of the mesh. Since the material is not self-supporting in sizes much over 12 in. diameter, it requires support bars at the point of location in the vessel. In most instances it is wise to also install hold-down bars across the top of the mesh in accordance with manufacturers' instructions as the material will tend to blow upward with a sudden surge or pulsation of vapor in the system. Many

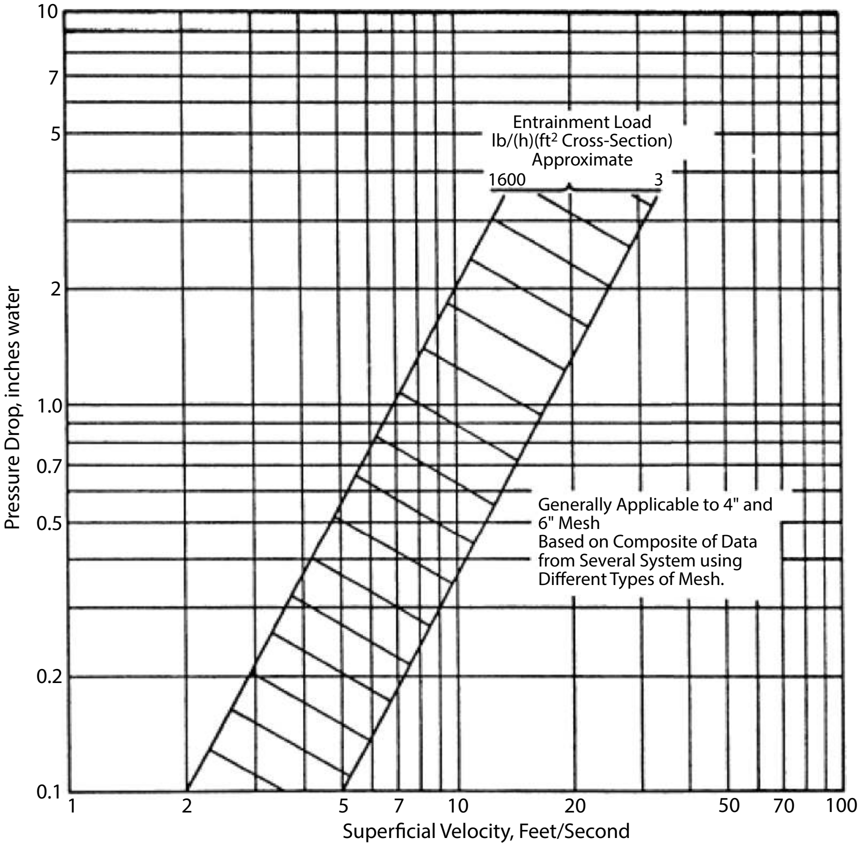


Figure 18.23 Typical pressure drop range for most wire mesh separators.

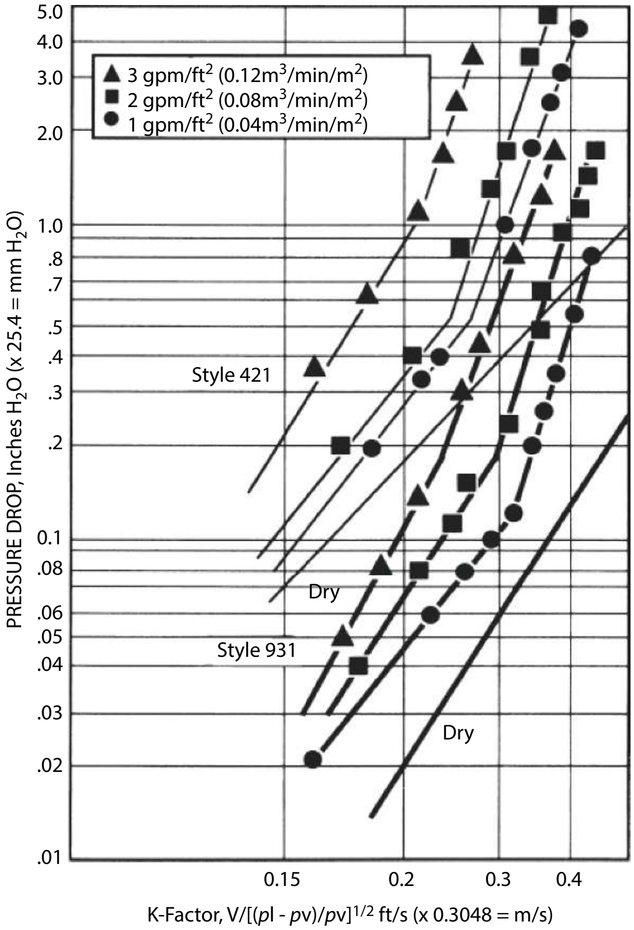


Figure 18.24 Typical wire mesh mist eliminator pressure drop curves for one style of mesh at three different liquid loadings. Others follow similar pressure drop patterns (by permission from Otto H. York Co. Inc.).

early installations made without the bars on top were soon found ineffective due to blowout holes, and wire particles were found in pipe and equipment downstream of the installation. Figures 18.25 and 18.26 show a typical installation arrangement in a vertical vessel. The mesh is wired to the bottom support bars and the hold-down on top.

A few typical arrangements of mesh in vessels of various configurations are shown in Figure 18.27. Note that in some units of Figure 18.27, the mesh diameter is smaller than the vessel. This is necessary for best operating efficiency under the system conditions, and applies particularly when using an existing vessel.

When placing mesh in small diameter vessels, it is important to discount the area taken up by the support ring before determining the operating velocity of the unit. For small 6-, 8-, and 12-in. vessels (such as in-line, pipe-with-mesh units) it is usual practice to use 6- or 8-in. thickness of mesh for peak performance.

Provide at least 6- to 12-in. minimum (preferably 18-in. minimum) disengaging space ahead of the inlet face of the mesh, that is, above any inlet nozzle bringing the liquid-carrying vapors to the vessel, or above any liquid surface held in the vessel. Leave 12-in. minimum of disengaging space above the mesh before the vapors enter the vessel vapor exit connection. The mesh may be installed in horizontal, vertical or slanting positions in circular, rectangular or spherical vessels. For locations where the liquid drains vertically through the mesh pad perpendicular or angular to its thickness dimension, care must be taken to keep velocities low and not to force or carry the liquid through to the downstream side of the mesh. Figures 18.28a and 18.28b illustrate typical wire mesh installations in vertical and horizontal vessels.

The check or specification form of Figure 18.30 is necessary and helpful when inquiring wire mesh entrainment units, either as the mesh alone, or as a complete turnkey unit including vessel.

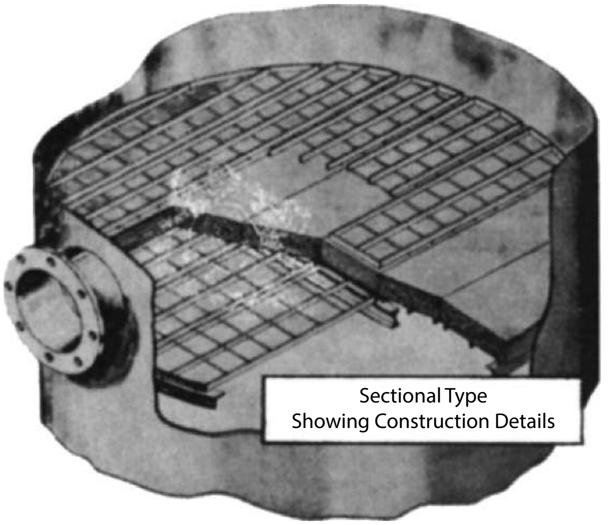


Figure 18.25 Typical installation of mesh strips in vertical vessel (courtesy of Otto H. York Co. Inc.).

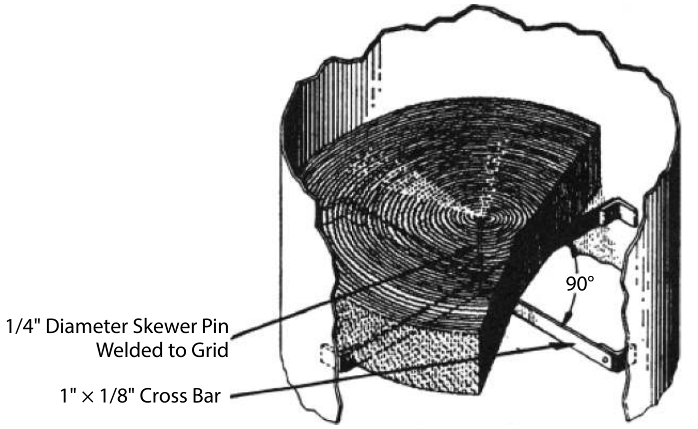


Figure 18.26 Typical installation of wound mesh pads in vertical vessel (courtesy of Metal Textile Corp. Bulletin ME 9-58).

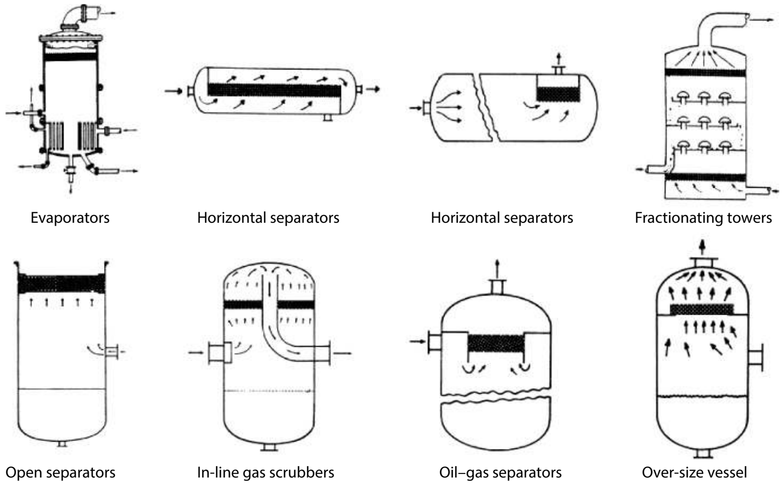


Figure 18.27 Typical mesh installation in process equipment (courtesy of Metal Textile Corp. Bulletin ME-7).

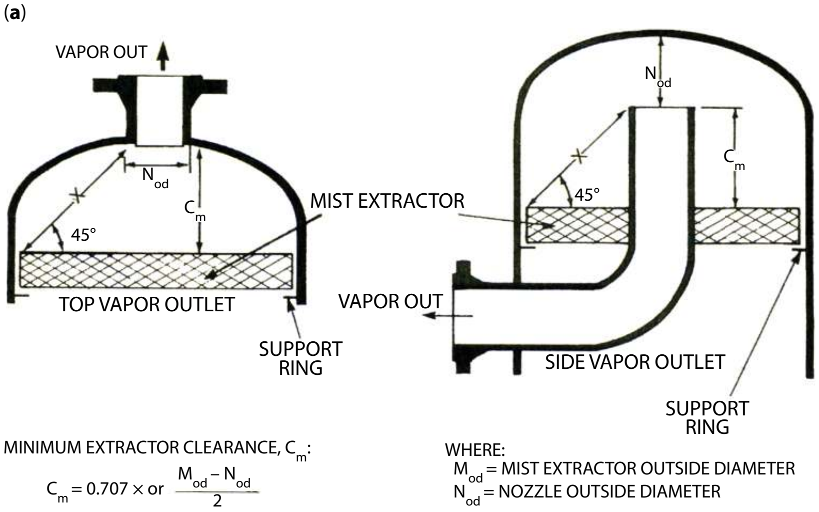


Figure 18.28a Minimum clearance-mesh type mist eliminators (source: GPSA Engineering Databook, 12th ed., section 7) [82].

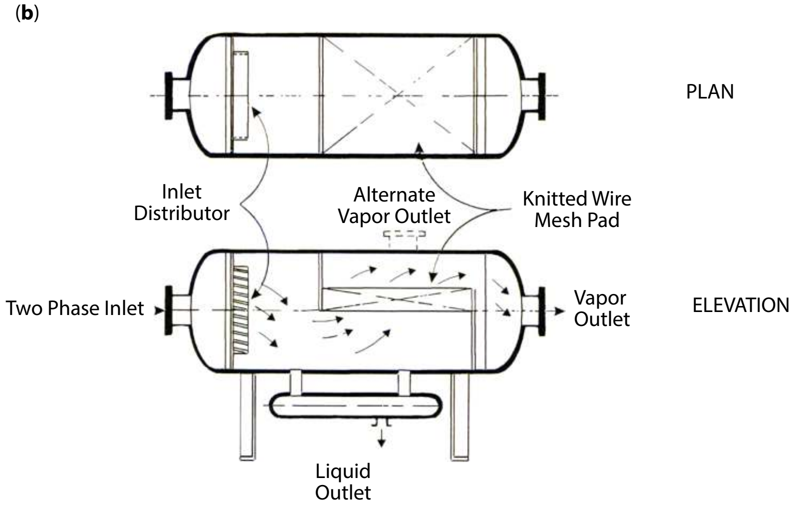
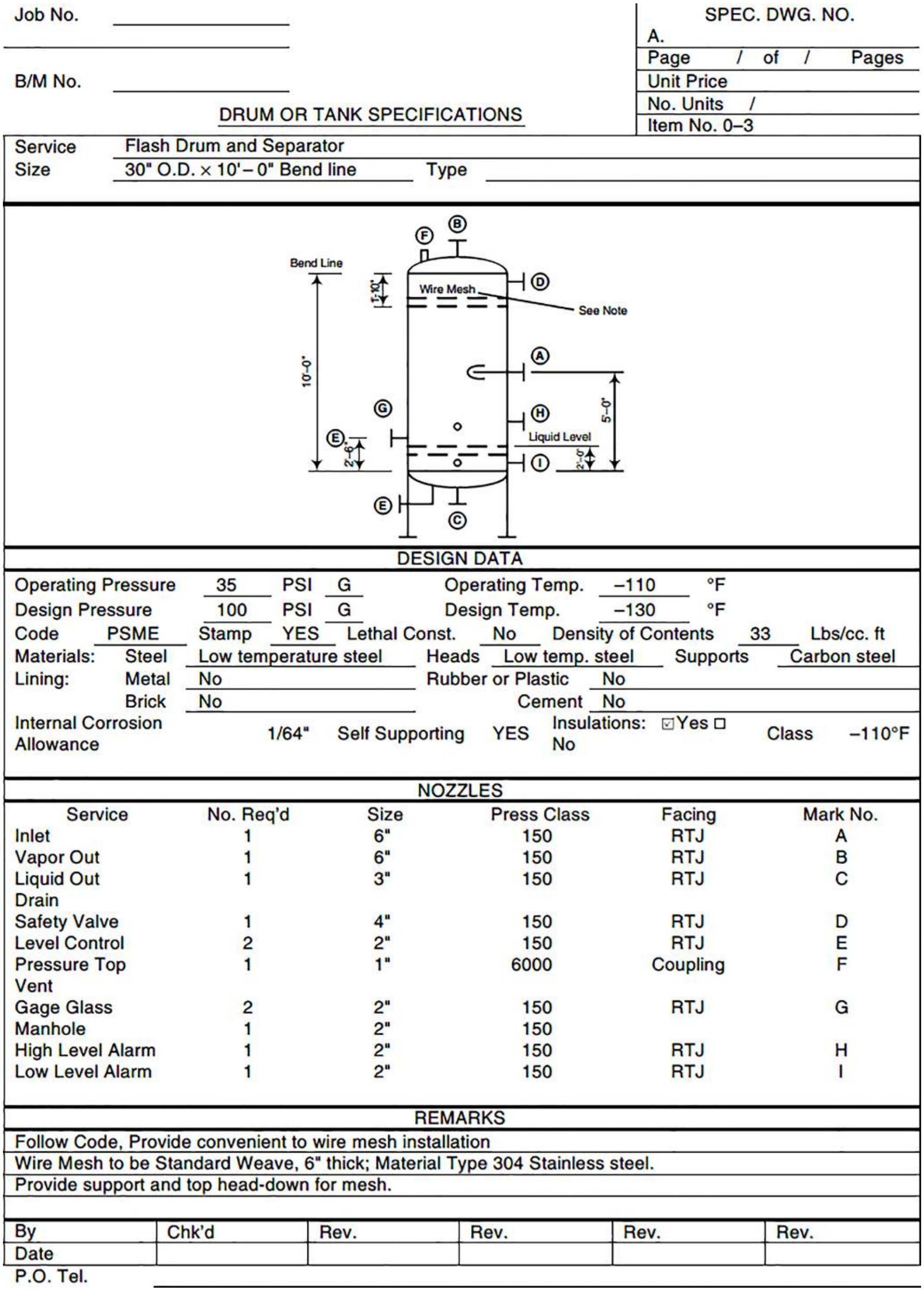


Figure 18.28b Horizontal separator with knitted wire mesh pad mist extractor and lower liquid barrel (source: GPSA Engineering Databook, 12th ed., section 7) [82].

For services where solids are present or evaporation of droplets on the mesh might leave a solid crust, it is usual practice to install sprays above or below the mesh to cover the unit with water (or suitable solvent) on scheduled (or necessary) operating times, as the plugging builds up. This is checked by a manometer or other differential pressure meter placed with taps on the topside and bottom side of the mesh installation.

A few case examples for guidance include the following:

- 2–3% caustic solution with 10% sodium carbonate. This condition might plug the wire mesh. Sprays would be recommended.
- Raw river water. This presents no plugging problem.
- Light hydrocarbon mist. This presents no plugging problem.
- Heavy oil with suspended matter. This might plug. A light oil or solvent spray would be recommended for flushing the mesh.



Job No.

B/M No.

SPEC. DWG. NO.
A.
Page / of / Pages
Unit Price
No. Units /
Item No. 0–3

DRUM OR TANK SPECIFICATIONS

Service Flash Drum and Separator
Size 30" O.D. × 10'– 0" Bend line Type

DESIGN DATA

Operating Pressure 35 PSI G Operating Temp. –110 °F
Design Pressure 100 PSI G Design Temp. –130 °F
Code PSME Stamp YES Lethal Const. No Density of Contents 33 Lbs/cc. ft
Materials: Steel Low temperature steel Heads Low temp. steel Supports Carbon steel
Lining: Metal No Rubber or Plastic No
Brick No Cement No
Internal Corrosion Allowance 1/64" Self Supporting YES Insulations: ☑Yes ☐ No Class –110°F

NOZZLES

Service	No. Req'd	Size	Press Class	Facing	Mark No.
Inlet	1	6"	150	RTJ	A
Vapor Out	1	6"	150	RTJ	B
Liquid Out	1	3"	150	RTJ	C
Drain					
Safety Valve	1	4"	150	RTJ	D
Level Control	2	2"	150	RTJ	E
Pressure Top	1	1"	6000	Coupling	F
Vent					
Gage Glass	2	2"	150	RTJ	G
Manhole	1	2"	150		
High Level Alarm	1	2"	150	RTJ	H
Low Level Alarm	1	2"	150	RTJ	I

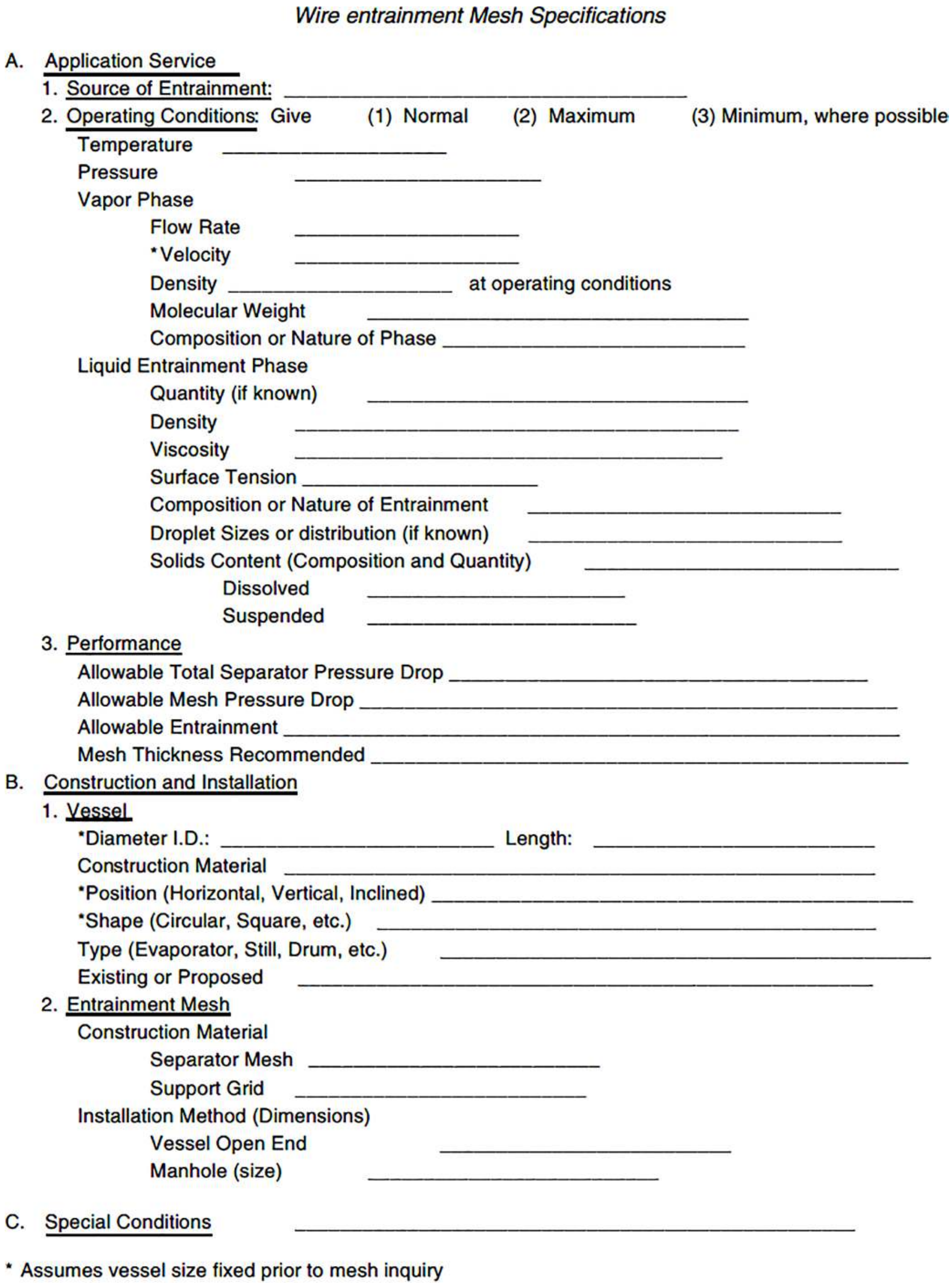
REMARKS

Follow Code, Provide convenient to wire mesh installation
Wire Mesh to be Standard Weave, 6" thick; Material Type 304 Stainless steel.
Provide support and top head-down for mesh.

By	Chk'd	Rev.	Rev.	Rev.	Rev.
Date					

P.O. Tel.

Figure 18.29 Specification design sheet for separator using a wire mesh.



Wire entrainment Mesh Specifications

A. Application Service
1. Source of Entrainment: ______________________________
2. Operating Conditions: Give (1) Normal (2) Maximum (3) Minimum, where possible
Temperature ________________
Pressure ________________
Vapor Phase
Flow Rate ________________
*Velocity ________________
Density ________________ at operating conditions
Molecular Weight ____________________________
Composition or Nature of Phase ______________________
Liquid Entrainment Phase
Quantity (if known) ____________________________
Density ________________________________
Viscosity ________________________________
Surface Tension __________________
Composition or Nature of Entrainment ________________________
Droplet Sizes or distribution (if known) ________________________
Solids Content (Composition and Quantity) ________________________
Dissolved ____________________
Suspended ____________________
3. Performance
Allowable Total Separator Pressure Drop ______________________________
Allowable Mesh Pressure Drop ____
Allowable Entrainment ____
Mesh Thickness Recommended ______________________________________
B. Construction and Installation
1. Vessel
*Diameter I.D.: ____________________ Length: ____________________
Construction Material ____
*Position (Horizontal, Vertical, Inclined) ____________________________________
*Shape (Circular, Square, etc.) ______________________________________
Type (Evaporator, Still, Drum, etc.) ____________________________________
Existing or Proposed ____
2. Entrainment Mesh
Construction Material
Separator Mesh ______________________
Support Grid ______________________
Installation Method (Dimensions)
Vessel Open End ______________________
Manhole (size) ______________________

C. Special Conditions ____

* Assumes vessel size fixed prior to mesh inquiry

Figure 18.30 Wire entrainment mesh specifications.

Example 18.5 Wire Mesh Entrainment Separator

Design a flash drum to separate liquid ethylene entrainment for the following conditions:

Volume of vapor = 465 ft^3/min. at −110°F and 35 psig
Density of vapor, $\rho_v = 0.30\ lb/ft^3$
Density of liquid ethylene, $\rho_L = 33\ lb/ft^3$

Allowable velocity for wire mesh:

$$V_a = k\sqrt{\frac{\rho_L - \rho_v}{\rho_v}}$$

Use, k = 0.35 for clean service, moderate liquid loading

$$\begin{aligned} V_a &= 0.35\sqrt{(33-0.3)/0.3} \\ &= 3.66\ \text{ft/s, allowable loading velocity} \end{aligned}$$

Use, $V_D = 0.75\ V_a$

Design velocity:

$$V_D = 0.75(3.66) = 2.74\ \text{ft/s}$$

Required vessel cross-section area:

$$A = 465/(60)\ (2.74) = 2.83\ \text{ft}^2$$

Vessel diameter:

$$D = \sqrt{\frac{2.83(4)}{\pi}} = 1.898 = 1' - 11''$$

Try: 2′– 0″ I.D. vessel

Deduct 4 in. from effective diameter for 2-in. support ring inside.

$$24'' - 4'' = 20''$$

Net area:

$$A = \frac{\pi(20)^2}{4(144)} = 2.18\ \text{ft}^2.$$

$$\text{Actual velocity at ring: } 2.74\left(\frac{2.83}{2.18}\right) = 3.56\ \text{ft/s.}$$

This is 97% of maximum allowable design, too high.

Second, try to increase diameter to next standard dimension, 2 ft, 6 in. Although intermediate diameters could have been selected, the heads normally available for such vessels run in 6-in. increments (either O.D. or I.D.).

Net inside diameter at support ring:

$$3''\text{O.D.} - 4'' - 3\Big/4'' = 25\frac{1}{4''}$$

Note that vessel wall assumed 3/8 in. thick. Net area = 3.46 ft^2.

$$\text{Actual velocity at ring: } \left(\frac{2.83}{3.46}\right)(2.74) = 2.24 \text{ ft/s}$$

Percent design velocity: 2.24(100)/3.66 = 61.3%. This is acceptable operating point.

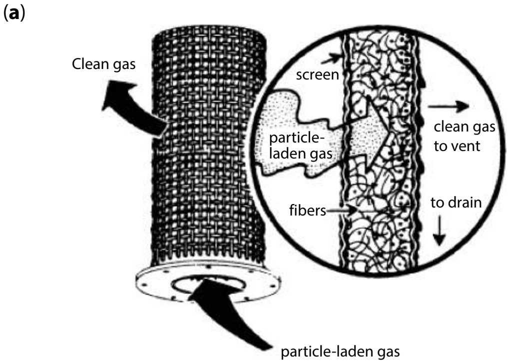
Note that if 28-in. O.D. × 3/8-in. wall pipe is available this could be used with weld cap ends, or dished heads. The percent design velocity would be = 71.8%. This is also an acceptable design.

Pressure drop is in the order of 0.1–0.5 in. H_2O.

Notes: Since this vessel will operate as a flash drum with a liquid level at approximately 1/4 of its height up from bottom, place the inlet at about center of vessel (Figure 18.29). Use stainless 304 mesh due to low temperature operation. Carbon steel is too brittle in wire form at this temperature.

Fiber Bed/Pad Impingement Eliminators

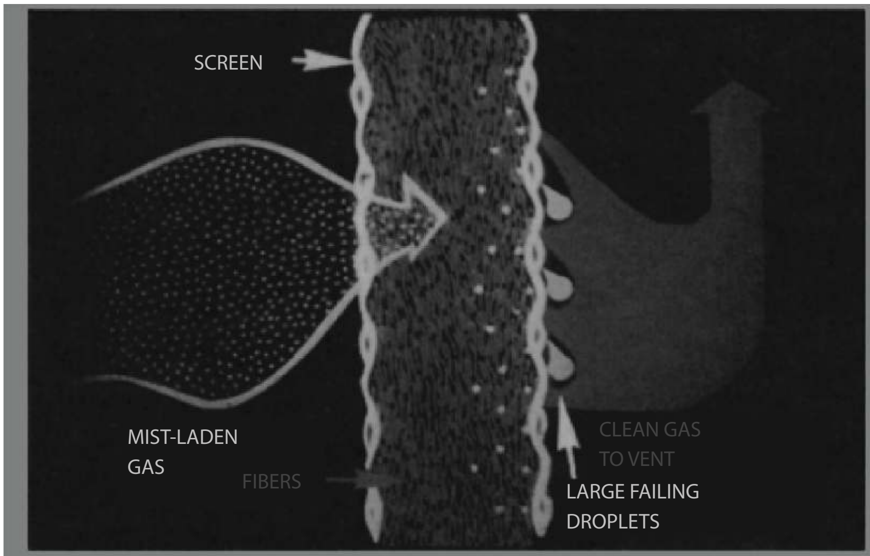
The use of fiber packing held between wire mesh containing screens is best applied in the very low micron range, generally 0.1 to >3 μm with recoveries of entrained liquid of up to 99.97% (by weight). Figures 18.31a–c illustrate the design concept and its corresponding data table indicates expected performance. The fibers used mean the bed packing can be fabricated from fine glass, polypropylene fibers, or can be selected to be the most resistant to the liquid mist entering the unit from corrosive plant operations such as sulfuric acid, chlorine, nitric acid, ammonia scrubbing for sulfur oxides control, and many others. The entrained particles are removed by direct interception, inertial impaction, and Brownian capture.



Materials of construction
Packing of York-Fiberbed high efficiency mist eliminators consists of ceramic, glass, polypropylene, fluoropolymer fibers. Cages and frames are fabricated from all stainless steels and other weldable alloys as well as FRP.

Figure 18.31a Details of a cylindrical York–Fibered® mist eliminator (courtesy of Otto H. York Co., Inc. Bulletin in 55B).

(**b**)



Available FLEXIFIBER

Mist Eliminator Styles

KOCH Type	Primary Collection Mechanism	Collection Efficiency		Element Pressure Drop (Inches WG)	Bed Velocity (Ft/Min)
		Particle Size (Microns)	Efficiency (%)		
BD	Brownian Diffusion	>3	Essentially 100	2–20	5–40
		<3	Up to 99,95		
IC	Impaction	>3	Essentially 100	7–9	250–450
		1–3	95–99+		
IP	Impaction	>3	Essentially 100	5–7	400–500
		1–3	85–97		
		0.5–1	50–85		
IS	Impaction	>3	Essentially 100	1–2	400–500
		<3	15–30		

NOTE: Since Types IC, IP and IS operate primarily by Impaction, the above collection efficiencies drop off at gas flows below 75% of design rates and depend on the speciifc gravity of the collected liquid.

Flexifiber® Type BD: Normally cylindrical in shape and avaiable in a wide variety of materials and sizes. Bulk packed elements are Mark I series.

Also available as wounded beds (Mark II or III) for lower initials cost and ease of repacking, and with equivalent collection efficiency at equal or lower pressure drop.

Flexifiber® Type IC: Normally cylindrical in shape and available in a variety of materials and sizes.

Flexifiber® Types IP and IS: Normally rectangular in shape and available in various metals.

Figure 18.31b Fiber-Pack® mist eliminator pack separators (by permission from Kock Engineering Co. Inc.).

(c)

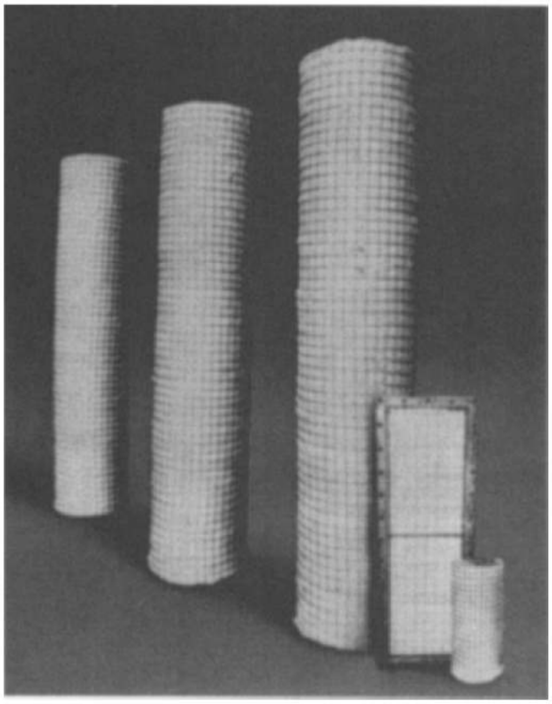


Figure 18.31c Typical fiberbed mist eliminators are available in both candle and panel configurations (by permission, Otto H. York Co. Inc.).

The design rating for this equipment is best selected by the manufacturers for each application.

The concept of removal of entrained liquid particle is essentially the same as for wire mesh designs, except the particle size removed may be much smaller. Just as for other types of mist eliminators, the performance is affected by the properties of the liquid particles, entraining gas, temperature, pressure, liquid viscosity, particle size distribution of entrained material and the quantity of total entrainment, and the desired process removal requirement. Some designs of these units provide excellent performance removal efficiencies at a wide range of rates (turndown), even at low gas rates.

Pressure drop is usually low depending on many factors, but can be expected in the range of 2–20 in. H_2O [32].

Baffle Type Impingement. There are many baffle type impingement separators. The efficiency of operation for entrainment is entirely a function of the contacting action inside the particular unit. There are no general performance equations which will predict performance for this type of unit; therefore, manufacturers' performance data and recommendations should be used. A few of the many available units are shown in Figures 18.32–18.35. Many use the Chevron-style vertical plates as shown in Figures 18.19a and 18.34.

Baffles (Chevron/Vanes). One of the common impingement plate assemblies is of the Chevron "zig-zag" style of Figures 18.19a and 18.34. This style of impact separation device will tolerate higher gas velocities, high liquid loading, viscous liquids, reasonable solids, relatively low-pressure drops. The collected liquid droplets run down on the plate surfaces counter-current to the up-flowing gas stream (McNulty [33] for performance study).

Spacing of the plates and their angles is a part of the design using the manufacturers' data. Multiple pass designs can result in higher recovery efficiencies. The units can be designed/installed for vertical or horizontal flow.

Some of the same physical properties of the liquid and gas phases as well as temperature and pressure and the amount of entrained liquids (or solids if present) and the expected particle size and its distribution control the design and performance of these units also.

Figure 18.32 Wall-wiping centrifugal type separator (courtesy of Wright Austin Co.).

For preliminary selection:

$$V_D = k\left[(\rho_l - \rho_v)/\rho_v\right]^{1/2} \tag{18.75}$$

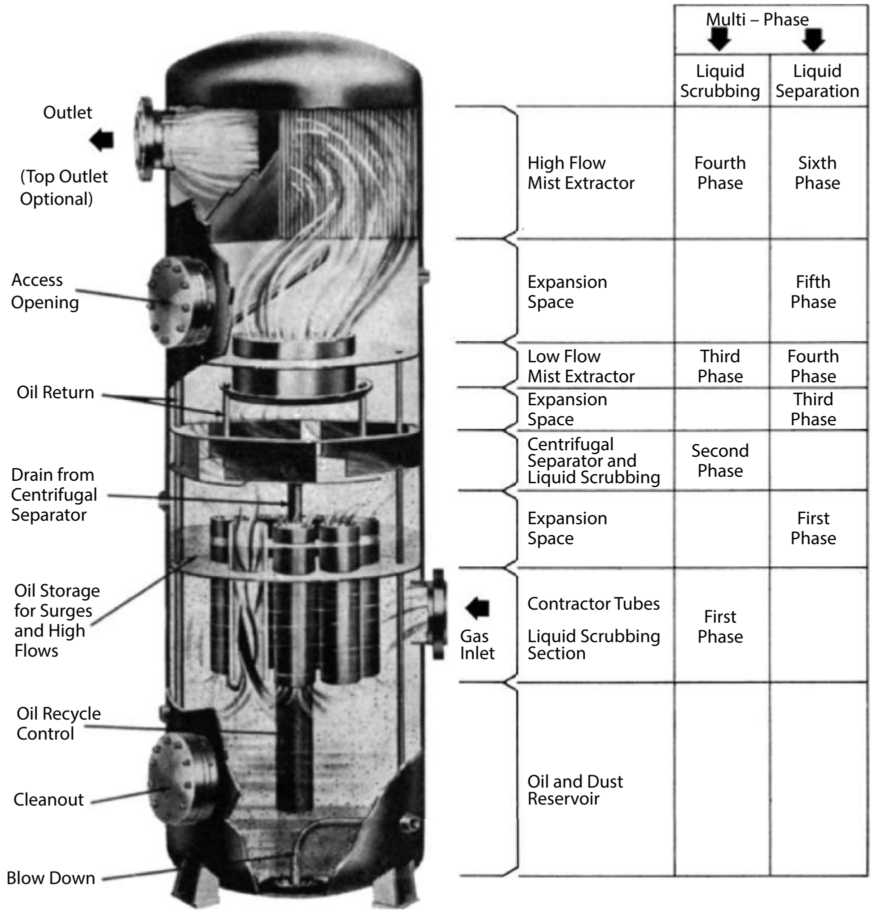
where

ρ_v = vapor density, lb/ft^3 at actual conditions
ρ_L = liquid density, lb/ft^3 at actual conditions
k = 0.40 for up-flow at 0.65 for horizontal flow.

For required flow area estimate only,

A = (ACFS)/V_D, ft^2
A = area ft^2
ACFS = actual flow, ft^3/s
V_D = design velocity, ft/s

Generally, this style of unit will remove particles of 12 to 15 μm efficiently. The typical droplet separator is shown for an air-water system in Figure 18.19a. This will vary for other systems with other physical properties. The variations



	Multi – Phase	
	Liquid Scrubbing	Liquid Separation
High Flow Mist Extractor	Fourth Phase	Sixth Phase
Expansion Space		Fifth Phase
Low Flow Mist Extractor	Third Phase	Fourth Phase
Expansion Space		Third Phase
Centrifugal Separator and Liquid Scrubbing	Second Phase	
Expansion Space		First Phase
Contractor Tubes Liquid Scrubbing Section	First Phase	
Oil and Dust Reservoir		

Figure 18.33 Multiphase gas cleaner (courtesy of Blow-Knox Co.).

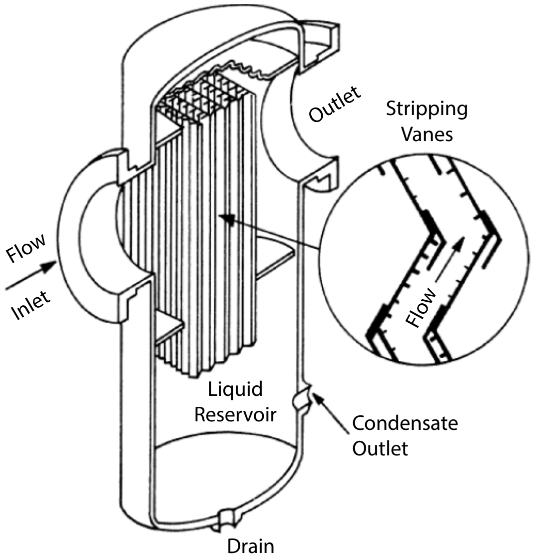


Figure 18.34 Impingement separator (courtesy of Peerless Manufacturing Co.).

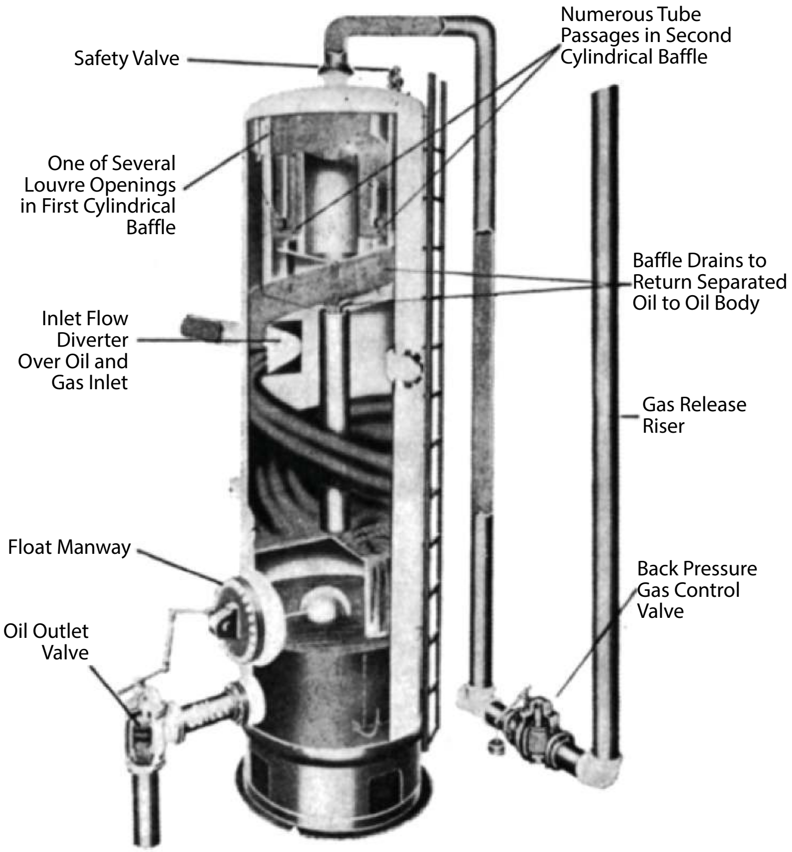


Figure 18.35 Combination separator (courtesy of National Tank Co.).

in capacity (turndown) handled by these units is in the range of 3–6 times the low to maximum flow, based on k values [33].

A liquid-liquid separator used for removing small, usually 2% or less, quantities of one immiscible liquid from another is termed “coalescer”. This unit is not gravity settlers, but agglomerates the smaller liquid by passing through a surface contact medium such as excelsior, hay, cotton or wool bats, or cartridges of fibers similar in nature and weave to those of Tables 18.12a and b. Figure 18.36 illustrates some of these types.

Efficiency. The efficiency of this type of unit varies, and is a function of the effectiveness of the impingement baffling arrangement. About 70% of separator applications can use the line-type unit; the other 30% require the vessel construction. The preference of the designer and problems of the plant operator are important in the final selection of a unit to fit a separation application. The efficiency for removal of liquid and solid suspended particles is 97–99%+ when handling 15 μm particles and larger. For steam service, a typical case would be 90% quality entering steam with 99.9% quality leaving.

Some units will maintain a reasonable efficiency of separation over a range of 60–120% of normal performance rating while other types will not. This flexibility is very peculiar to the internal design of the unit. Some units are guaranteed to reduce mechanical entrainment loss to less than 0.1 gal. per million standard cubic feet of entraining gas.

Pressure Drop. Pressure drop in most units of this general design is very low, being in the order of 0.1–3 psi.

How to Specify. Manufacturers’ catalogs are usually available and complete with capacity tables for the selection of a unit size. However, it is good practice to have this selection checked by the manufacturer whenever conditions will allow. This avoids misunderstandings and misinterpretations of the catalog, thereby assuring a better selection for the separation operation, and at the same time the experience of the manufacturer can be used to advantage.

Table 18.12a Partial list of commercial crossflow microfilter media-materials and geometries.

Material	**Geometries**					
	Pleated sheet	**Tubular**	**Spiral wound**	**Tubular MC***	**Hollow fiber**	**Flat sheet**
1. Polymers						
Cellulosics	⬣		⬣		⬣	⬣
Polysulfone		⬣	⬣		⬣	⬣
Polyvinylidene fluoride		⬣	⬣			
Acrylic					⬣	
Polytetrafluoroethylene		⬣				⬣
Polybenzimadazole			⬣			⬣
Polypropylene					⬣	⬣
Nylon	⬣				⬣	⬣
2. Ceramics						
Alumina				⬣		⬣
Zirconia/alumina				⬣		
Zirconia/sintered metal		⬣				
Zirconia/carbon		⬣				
Silica		⬣				
Silicon carbide				⬣		
3. Sintered metal						
Type 316 stainless steel		⬣				
Other alloys		⬣				

*MC = Multi-channel monolithic elements
Source: By permission from Michaels [68].

Table 18.12b Partial list of crossflow microfilter media in chemical service applications.

Chemicals	Compatible media												
Table II	**Ceramics**	**Acrylics**	**Cartoon**	**Poly-benzimadazole**	**Polysulfone**	**Polytetra-fluoroethylene**	**Polypropylene**	**Polyvinylidene fluoride**	**Other polymers**	**Sintered halide-resistant alloys**	**Sintered 316 stainless steel**	**Other sintered media**	**Sintered chloride-resistant alloys**
Alkanes, alkenes, and aromatic hydrocarbon, below 100°C	●		●			●				●	●	●	●
Oxygen-containing organics, below 100°C	●		●			●	●	●		●	●	●	●
Chlorinated organics, below 100°C	●		●			●	●			●	●	●	●
Esters, below 100°C	●		●			●	●	●		●	●	●	●
Organics at 100–200°C	●		●			●				●	●	●	●
Organics at 200–600°C	●		●							●	●	●	●
Organics at 600–900°C	●		●							●	●	●	●
Gases:													
Inert, or low reactivity	●		●	●	●	●	●	●		●	●	●	●
Oxygen	●												
Chlorine	●		●			●							●
Other reactive gases	●					●							

(*Continued*)

Table 18.12b Partial list of crossflow microfilter media in chemical service applications. (*Continued*)

Chemicals	**Compatible media**												
Table II	**Ceramics**	**Acrylics**	**Cartoon**	**Poly-benzimadazole**	**Polysulfone**	**Polytetra-fluoroethylene**	**Polypropylene**	**Polyvinylidene fluoride**	**Other polymers**	**Sintered halide-resistant alloys**	**Sintered 316 stainless steel**	**Other sintered media**	**Sintered chloride-resistant alloys**
Aqueous solutions:													
pH = 3–7, no chlorides	●		●		●	●	●	●	●		●		
pH = 7–10, no chlorides	●		●		●	●	●	●	●		●		
pH = 0–3 (except HF)	●		●	●	●	●	●	●				●	
pH = 3–10, chloride present	●		●		●	●	●	●	●				●
HF, with pH < 3	●		●			●	●	●		●			
pH = 10–13		●	●	●	●	●	●	●		●	●	●	●
pH > 13			●			●	●	●		●	●	●	●
Concentrated acids	●		●										
Steam (>100°C)	●		●			●		●		●	●	●	●
Oxidants (e.g., bleach)	●		●		●	●	●	●		●	●	●	●

Source: By permission from Michaels [68].

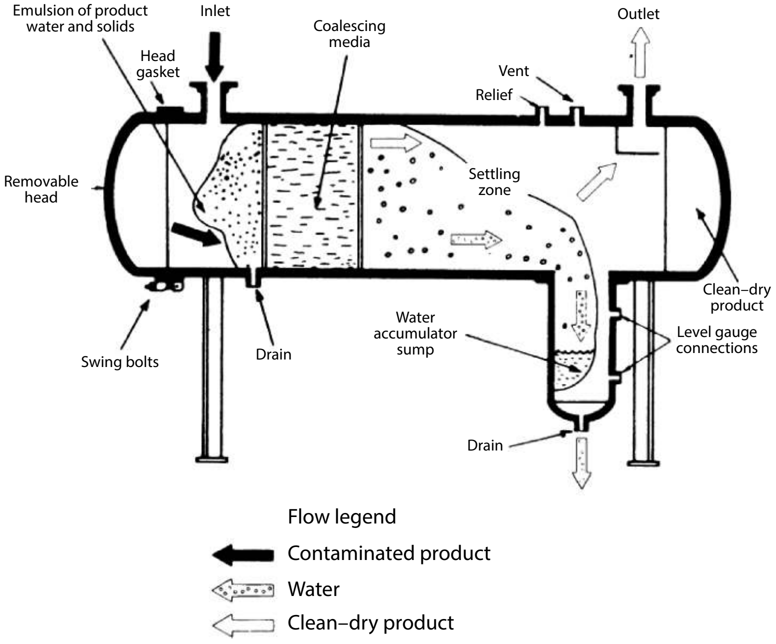


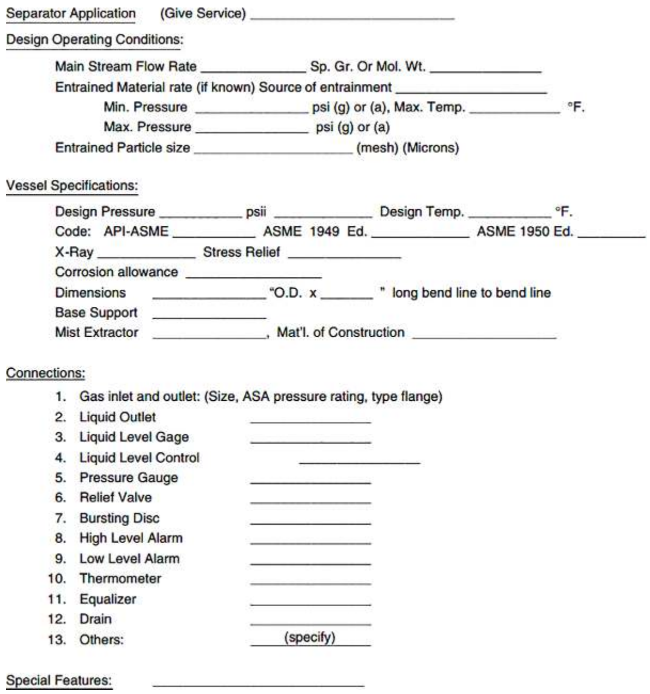
Figure 18.36 Typical coalescer unit (by permission from Facet Enterprises Inc., Industrial Div.).

With a manufacturer's catalog available, do the following:

1. Establish the minimum, normal and maximum gas flow for the system where the unit will operate. This is usually in standard cubic feet per minute, per hour or per day. Note the catalog units carefully and also that the reference standard temperature is usually 60°F for gas or vapor flow.
2. Use the rating selection charts or tables as per catalog instructions. For specific gravity or molecular weight different than the charts or tables, a correction factor is usually designated and should be used.
3. Note that some units use pipe line size for the separator size designation, others do not.
4. From the system operating pressure, establish the pressure rating designation for the separator selection.
5. Note that most separators for pressure system operations are fabricated according to the American Society of Mechanical Engineers (ASME) code.
6. Specify special features and materials of construction, such as alloy or nonferrous impingement parts, or entire vessel if affected by process vapor and liquid. Specify special liquid reservoir at the base of unit if necessary, for system operations. Line units normally have dump traps or liquid outlet of separator, while vessel type often uses some type of liquid level control.
7. Prepare a specification sheet (Figure 18.37).

Note that these units should not be connected in lines larger than their pipe inlet, since inlet velocity conditions are very important, the swaging down or reduction tends to produce a jet effect by the gas upon the mist eliminator unit. This may erode the unit and cause other erratic performance.

Dry-Packed Impingement Beds. Although this type of unit is not used as frequently as most of the others discussed, it does have some specific applications in sulfuric acid mist removal and similar very difficult applications. The unit consists of a bed of granular particles or ceramic packing, sometimes graduated in size, operating dry as far as external liquid application to aid in the separation. The superficial velocities of 0.5 to 8 ft/s. through the unit are rather low for most separators therefore the vessels become large. Due to the packed heights of 2 ft (minimum) and higher, the pressure drop can be appreciable. Particle removal may be 0.5–5 μm at 99% efficiency for a good design. These units will plug on dust service and must be back washed to regain operability at reasonable pressure drops.



Separator Application (Give Service) ____________

Design Operating Conditions:

Main Stream Flow Rate ________ Sp. Gr. Or Mol. Wt. ________
Entrained Material rate (if known) Source of entrainment ________
Min. Pressure ________ psi (g) or (a), Max. Temp. ________ °F.
Max. Pressure ________ psi (g) or (a)
Entrained Particle size ________ (mesh) (Microns)

Vessel Specifications:

Design Pressure ________ psii ________ Design Temp. ________ °F.
Code: API-ASME ________ ASME 1949 Ed. ________ ASME 1950 Ed. ________
X-Ray ________ Stress Relief ________
Corrosion allowance ________
Dimensions ________ "O.D. x ________ " long bend line to bend line
Base Support ________
Mist Extractor ________, Mat'l. of Construction ________

Connections:

1. Gas inlet and outlet: (Size, ASA pressure rating, type flange)
2. Liquid Outlet ________
3. Liquid Level Gage ________
4. Liquid Level Control ________
5. Pressure Gauge ________
6. Relief Valve ________
7. Bursting Disc ________
8. High Level Alarm ________
9. Low Level Alarm ________
10. Thermometer ________
11. Equalizer ________
12. Drain ________
13. Others: (specify)

Special Features: ____________

Figure 18.37 Baffle type separator specifications.

18.12 Centrifugal Separators

Centrifugal separators utilize centrifugal action for the separation of materials of different densities and phases. They are built in (a) stationary and (b) rotary types. Various modifications of stationary units are used more than any other kind for separation problems. The cost is moderate; it is simple in construction, and is reasonably flexible in service, being useful for gas–liquid or gas–solid systems. In addition to serving as finishing separators centrifugal units are also used to take a "rough cut" into a separation problem. They may be followed by some additional unit of special cyclone action or filtration through woven cloth pads, and so on, to completely remove last traces of entrained particles.

Stationary Vane

The stationary vane type is quite popular and is adapted to many applications. It is used in vessels or pipe lines as illustrated in Figures 18.38–18.41d. They are usually of high efficiency for both liquid and solid particles such as rust, scale, dirt, and so on. When the system is dry with dust a special design is used.

Efficiency

The efficiency of centrifugal units is:

Type	Efficiency range
High Velocity	99% or higher, of entering liquid
Stationary Vanes	Residual entrainment 1 ppm or less
Cyclone	70–85% for 10 μm, 99% for 40 μm and larger. For high entrainment, efficiency increases with concentration
Rotary	98% for agglomerating particles

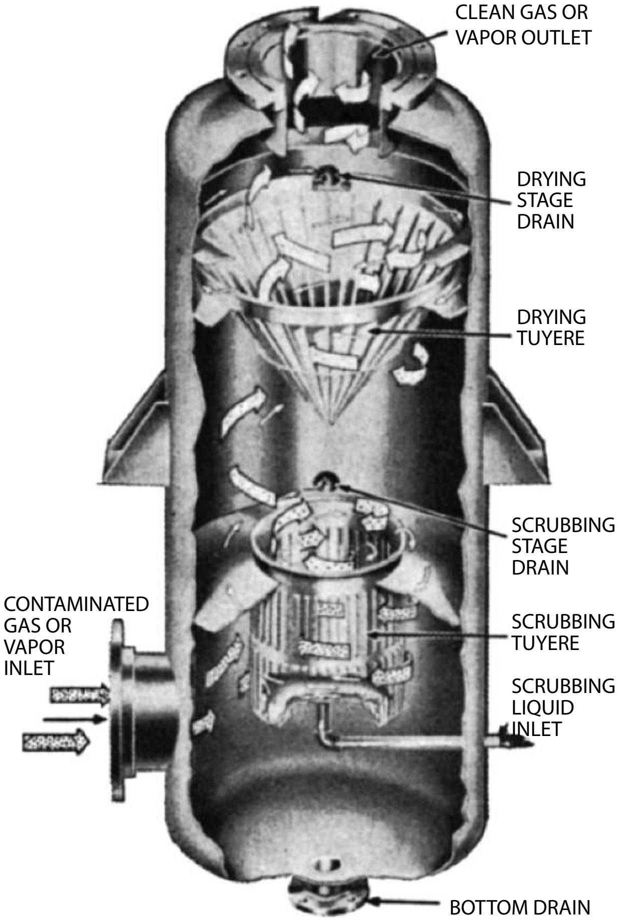


Figure 18.38 Scrubber with internal liquid feed (courtesy of Centrifix Corp.).

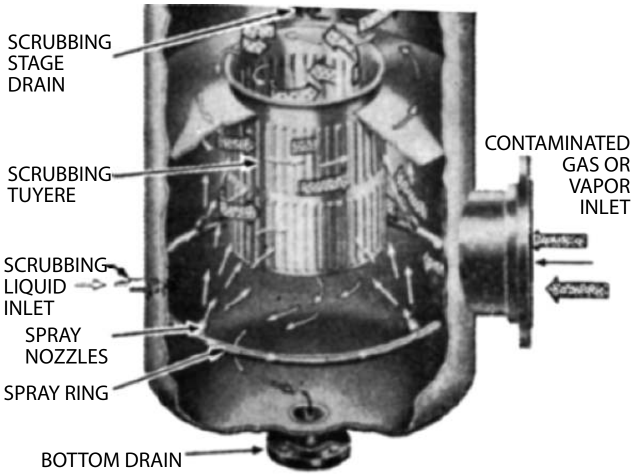


Figure 18.39 Scrubber with spray ring as alternate arrangement for Figure 18.38 (courtesy of Centrifix Corp.).

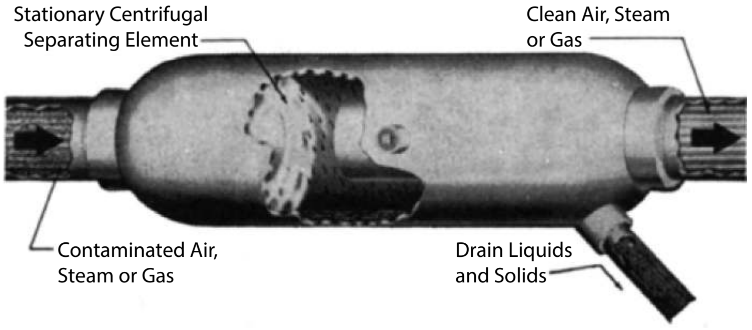


Figure 18.40 Line-type centrifugal separator (courtesy of V.D. Anderson Co.).

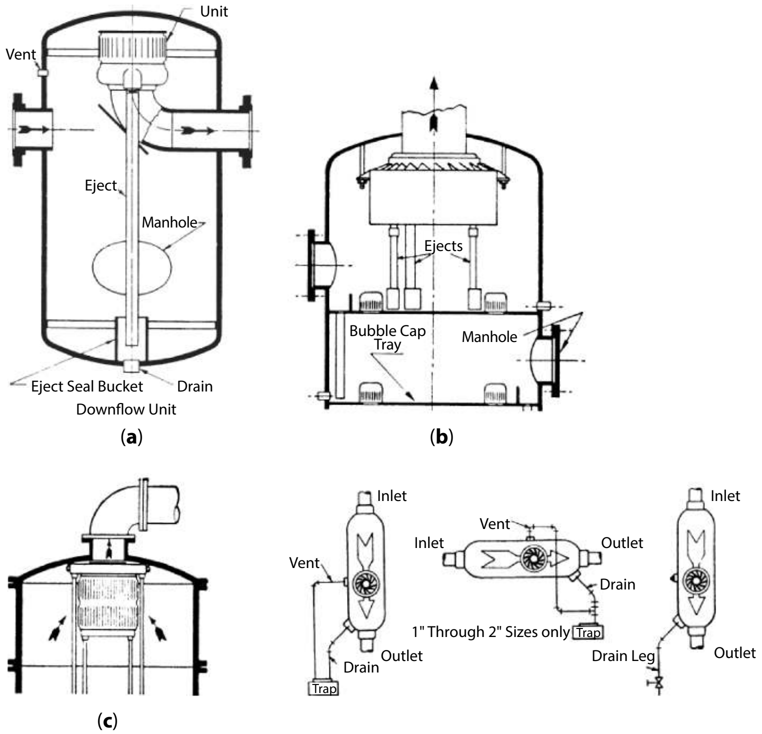


Figure 18.41 Centrifugal separator applications (by permission from Centrifix Corp.).

Two-Phase Separators

A two-phase separator for separating gas and liquid must consist of the following elements:

- A separation section to remove bulk liquid from gas.
- Sufficient liquid capacity to handle surges from the headers.
- Mist eliminator
- Level control.

The separator can be further classified into scrubbers and separators. A scrubber is a vessel designed to handle streams with high gas-to-liquid ratios. The liquid is generally entrained as mist in the gas or is free-flowing along the pipe wall. These vessels usually have a small liquid collection section. It is not an elaborate design and does not have enough liquid capacity to handle surges. Scrubbers handle only small quantities of liquid compared to a separator. Separators usually have facilities to remove liquid particles up to 10 μm. Particles less than 10 μm are referred to as particulates and are not handled in separators.

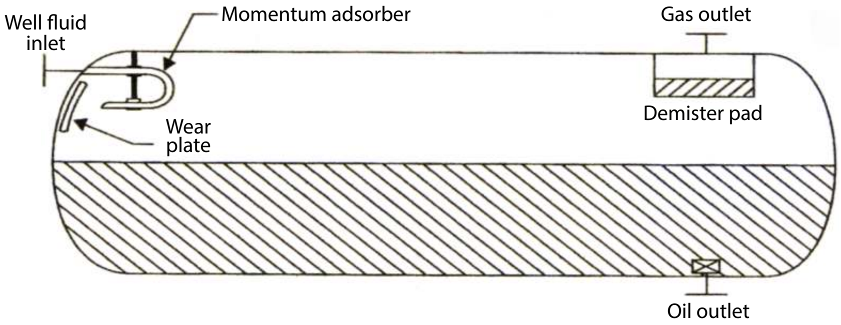


Figure 18.42 Two-phase separator [77].

Figure 18.42 shows a typical two-phase separator. The inlet stream enters an inlet distributor or an inlet momentum device, where particles of liquid more than 500 μm are removed. The larger particles are thrown into the target surface by centrifugal action caused by a change in the direction of the fluid. The particles coalesce on this surface and fall down to the bottom of the vessel or fall to the liquid layer. The gas then passes through the gravity separation section. Here, particles of diameter more than a definite size (100–150 μm) are removed by the action of gravity. Finally, the gas passes through a demister pad where particles more than a definite size of 10μ are removed by coalescing into large particles and fall into the liquid layer.

The retention time of oil in the separator should be more than the minimum values recommended by API 12J standard [81].

Oil relative density	Residence time (min)
Below 0.85	1
0.85–0.93	1–2
0.93–1.0	2–4

The residence time as shown can be more depending upon the presence of emulsions, foaming density, and so on. Since level control is provided for liquid, a certain amount of liquid volume should be provided for effective level control.

Vessel Internals

The role of the vessel internals is to assist in coalescing droplets to a diameter large enough for separation to occur (e.g., 150 μm size). The larger the droplets, the greater the internal velocity and the better the separation. There are a number of different internals such as mesh pads, vane packs and plate packs that can be selected depending on fluid properties, flow paths and location within the vessel. Internals coalesce and separate small droplets by making the fluid path as arduous as possible in order to collide and combine smaller droplets, forming larger droplets, giving better and easier separation. Internals are considered in more detailed design stages and the design of internals should be carried out in collaboration with suppliers and computational fluid dynamic (CFD) analysis in order to ensure that the most optimum setup is achieved. Figure 18.43 shows a photograph of the separator internals and Figure 18.44 shows the reflux vessel configuration with nozzles and instrument connections.

The vessel internals can significantly affect the operating performance of an oil/gas separator through the following:

- Flow distribution
- Drop/bubble shearing and coalescence
- Foam creation
- Mixing
- Level control

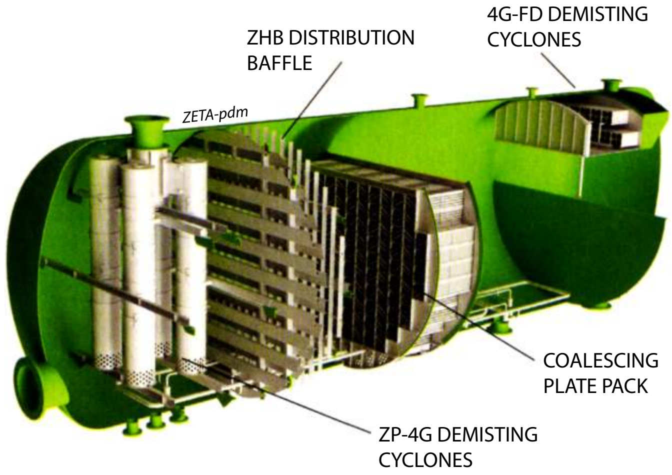
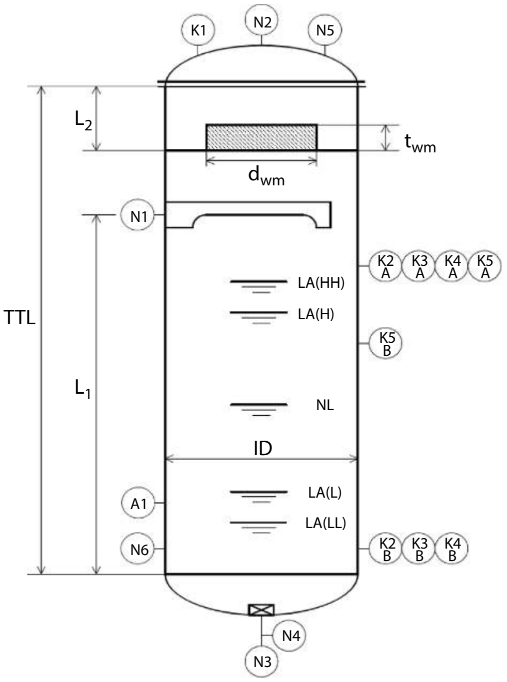


Figure 18.43 Internals of a separator [84].



Main Dimensions

TTL = 2720 mm
ID = 800 mm
L_1 = 1490 mm
L_2 = 250 mm
t_{wm} = 100 mm
d_{wm} = 400 mm

Level setting relative to bottom TL

LA(HH) = 1165 mm
LA(H) = 790 mm
LA(L) = 425 mm
LA(LL) = 300 mm

General Remark

(1) Vessel shall be designed in accordance to DEP 31.22.05.11-Gen
(2) Product outlet nozzle N3 to be equipped with a type A vortex breaker (see standard drawing S 10.010-F).
(3) The LA(HH) and LA(LL) are level settings based on recommended intervention times. Engineering contractor shall decide whether the level alarms shall be changed to LZA settings based on process safeguarding.
(4) As feed inlet a half-open pipe is installed. The opening is directed downwards.
(5) Refer to standard drawings 20.028/029/030 for wiremesh support details.
(6) The demister void fraction shall be >97% and the wire thickness between 0.23 and 0.28 mm.
(7) The Shell of the feed vessel is made of "clean" carbon steel ("Z" quality for wet H2S service) lined with GRE and extra corrosion allowance (6 mm).
(8) Internals (half-open pipe and mistmat) are made of Stainless Steel 316L.

Figure 18.44 Reflux vessel configurations with nozzles and instrument connections [82].

When the pressure is reduced on certain types of crude oil, tiny bubbles of gas are encased in a thin film of oil when the gas exits the solution. This may result in foam, or froth being dispersed in the oil and creates what is known as "foaming" oil. In other types of crude oil, the viscosity and surface tension of the oil may mechanically lock gas in the oil and can cause an effect similar to foam. Oil foam is unstable or long-lasting unless a foaming agent is present in the oil. An agent are organic acids and high-gravity oils and condensates typically do not result in foaming situations.

Foaming greatly reduces the capacity of oil/gas separators because a much longer retention time is required to adequately separate a given quantity of foaming crude oil. Foaming crude oil cannot be measured accurately with positive-displacement meters or with conventional volumetric metering vessels. These problems, combined with the potential loss of oil/gas because of improper separation, emphasize the need for special equipment and procedures in handling foaming crude oil.

The main factors that aid in breaking foaming oil are:

- Settling
- Agitation (baffling)
- Centrifugal force

These factors of reducing or breaking foaming oil are also used to remove entrained gas from oil. Many different designs of separators for handling foaming crude oil have evolved. They are available from various manufacturers and some as standard foam handling units and some designed especially for a specific application. Silicone and fluorosilicone-based chemical defoamer are typically used in conjunction with cyclonic inlets to break foam. The chemical defoamer concentration is generally in the range of 5–10 ppm. For many Gulf of Mexico (GOM) crudes, 50–100 ppm is common.

Residence Times

After determining a vapor area, estimates for diameter, length and height will be known, which allows other level settings to be established and liquid residence time criteria to be checked. Liquid residence time is the time that liquid stays in the vessel. Generally, control times required for level control are sufficient to allow degassing to occur; however, additional time may be required for heavier or more viscous liquids such as crude oil as indicated in the Table below.

Oil gravity	Required residence time for degassing NLL vessel bottom (min)
≥35°API	3–5
20–30°API	1–2
10–20°API	2–4

Residence times for liquid–liquid separation are generally the limiting factor in selecting level settings. Effective volume for separation within a vessel is limited to the volume where the two-phases are in contact, once liquid leaves this section (e.g., oil holdup downstream of the weir), it cannot be considered as part of the retention volume. Therefore, the following rules should be applied:

- Residence time between the overall liquid level bands should be calculated based on total liquid (i.e., oil and water) flow.
- For both oil and water phases, the volume upstream of the weir, plus the volume in a single head can be considered.

There are two considerations when specifying residence time:

- Oil settling time to allow adequate water removal from the oil. This applies between NLL to iNLL
- Water settling time, to allow adequate oil removal from the water. This applies between iNLL and vessel bottom.

By entering initial estimates for level settings, the volume can be determined and using the feed flow rates the residence time can be calculated. For all cases, the residence time basis to be used can be obtained based on laboratory tests [GPSA]. Industry standard recommended residence times for the oil phase are illustrated below for guidance. Initially, oil and water should be given equal residence times in the vessel.

Calculations are carried out to establish whether the droplet will reach the interface layer in the time that will be taken for the travel using settling/buoyancy phenomena. Calculations are cross checked using certain minimum residence times required for oil and water to achieve the required separation. API 12J standard recommends the following residence times based on the oil gravity [81].

Oil gravity	**Required residence time for liquid/liquid separation (min)**
Above 35°API	3–5
Below 35°API	
≥37.8°C (100°F)	5–10
≥26.7°C (80°F)	10–20
≥15.6°C (60°F)	20–30

Selection of Separators

It is important to select the correct geometry and configurations for the design of separators. Table 18.13 provides functions of the vertical, horizontal, and spherical separators. The table further provides information on factors such as separation efficiency, maintenance, cost and so on. The criteria and rules for selection of a separator are as follows [87]. Figure 18.45 shows a block diagram in determining the correct configuration of a separator.

Step 1.	What is the purpose of the equipment?	Different types of separators are employed for different purposes. Table 18.13a is used as a guide for selecting the correct equipment.
Step 2.	What will be the application of the separator?	This could be low temperature separators, test separators or production separator.
Step 3.	What are the properties of the fluid? Are there any impurities?	The designer considers the parameters such as temperature, pressure, density and viscosity of the stream. The droplet size to be removed or retention.
Step 4.	How many phases will be separated?	This depends on stream properties. The separator could be a two- or three-phase separating gas from liquid or separating gas, oil and water.
Step 5.	What is the composition of the well-stream?	A horizontal separator should be used for high to medium gas–oil ratio streams while a vertical separator should be used for streams with low gas–oil ratio and when sand, paraffin or wax are present.
Step 6.	Are there any restrictions such as space or cost?	Horizontal separators are less expensive however, it occupies more space whereas vertical separators cost less and occupy less space.

Table 18.13 Summary of phase separators [81].

Function	**Vertical**	**Horizontal**	**Spherical**
Usage	Vertical separators are used when the gas–oil ratio is low, when sand, paraffin, or wax are present and when ease of level control is desired	This configuration is used for streams with high to medium gas–oil ratio and for flow streams with large volume of gas and/or liquids	Spherical separators are small leases operating at moderate pressure and they are used for low or intermediate gas–oil ratio
Cleaning	Easier to clean than horizontal vessels	Most difficult to clean because sludge have longer length to settle	Much easier to clean than the other configurations
Gas/Liquid Capacity	Handle streams with large liquid capacity	Handle streams with large gas capacity	Poor gas/liquid capacity
Surge space	Good surge space	Limited surge space	Least surge space
Separating efficiency	Less efficient	Best	Least efficient
Handling of foaming oil	Vertical separators have a less tendency to handle foaming oil	The best separator to handle foaming oil	Performs very poorly with foaming oil
Liquid level control	Liquid level control is not critical	Liquid level control is critical	Liquid level control is very critical
Maintenance and Inspection	Difficult associated with height hazards	Horizontal separators are much more accessible	Average to maintain
Installation	Difficult to install due to height and requires stronger mounting base	Average and not as difficult as vertical separators	Easy to install
Cost	Expensive to build	Less expensive as compared to the other configurations	The most expensive to build
Advantages	Vertical separators save space uses; their full diameter for gas flow at the top and have less tendency for liquid re-vaporization into the gas phase	Horizontal separators require less piping for field connections and can handle foaming crude as it reduces turbulence and foaming. Also space requirements can be minimized as several separators can be stacked together	Spherical separators have the advantage of easy shipping and installation due to its compact size
Disadvantages	Vertical separators require bigger diameter separator for a given gas capacity, which results in high capital and transportation cost	Liquid level control is more critical in horizontal separators compared to vertical and are more difficult to clean when wax, paraffin, and sand are deposited at the bottom of the separator	Spherical separators are expensive to build and very difficult to use especially for three-phase separation due to having very limited liquid settling section

Table 18.13a Different types of separators and their purposes [87].

Type of separator	**Purpose**
Phase Separator	Separation of mixed stream coming directly from an oil or gas well
Scrubber	This is designed from streams with high gas/liquid ratios for the separation of small liquid droplets from a gas phase
Free water knockout	Separation of hydrocarbon phases from a high water cut stream
Flash Chamber	Used for subsequent phase separation to process liquid hydrocarbons flashed from a separator

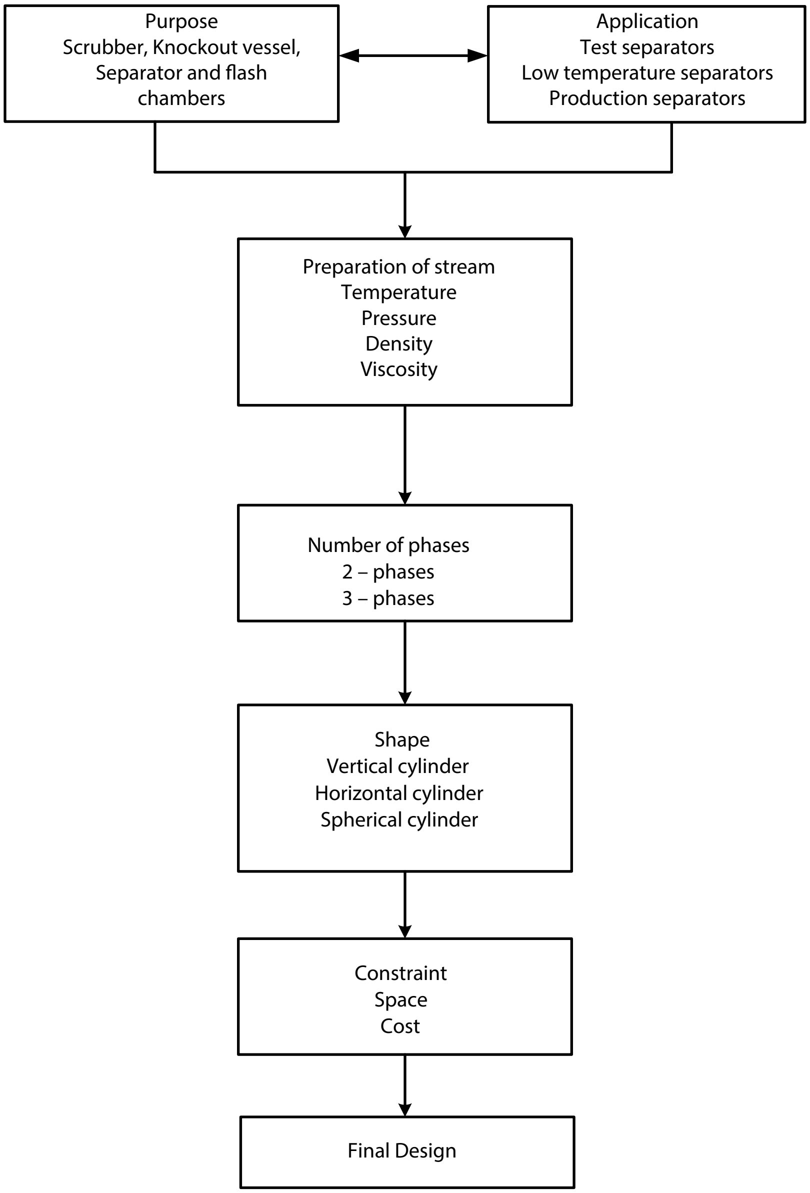


Figure 18.45 A block diagram for selection criteria of a separator [81].

Troubleshooting Gas–Liquid Separators

A common problem in oil and gas processing facilities is under performing vapor/liquid separators. The most common types of gas–liquid separators are:

Conventional separators (Vertical/Horizontal).

Scrubbers (i.e., Compressor Suction Scrubbers).

Slug Catchers (Vessel/Finger-type).

Gas "polishers" (Coalescing Filters/Filter Separators).

Low performance of separators results from:

- Selection of the wrong type of equipment for the application.
- Inadequate sizing methodology, although the correct equipment type is selected.

The type of separator required for an application depends largely on the gas–liquid ratio of the streams to be treated, and the flow variability of the process (Figure 18.46). As the flow variability increases with low to moderate gas–liquid ratios, the separator selection will move from a conventional separator to a slug catcher. For applications where there is a high gas–liquid ratio (i.e., very low liquid content), and the flow variability is moderately low, scrubbers and gas polishers would be the appropriate equipment selection depending upon the gas quality requirement for the treated stream. Common problems in separator performance are either too much liquid in the separated gas or inadequate slug/surge capacity and too much gas in the separated liquid. Figure 18.47 illustrates a troubleshooting methodology flow chart, and the problem here is too much liquid in the separated gas stream. In order to adequately troubleshoot separator performance, it is essential to understand the metrics of good performance, and the functions and analysis of the various components of the separation equipment. Table 18.14 provides a typical performance metrics. The specific performance requirements for a given separator are set by the sensitivity of the downstream process/equipment to the presence of liquids.

Gas–Liquid Separators

The process engineer is often required to design separators or knockout drums for removing liquids from process gas streams. The vessels used for processing in the chemical process industry (CPI) are principally of two kinds: those that are without internals and those with internals. Empty separators are drums that provide intermediate storage or surge of a process stream for a limited or extended period. Alternatively, they provide phase separation by settling.

The second category comprises equipment such as reactors, mixers, distillation columns, and heat exchangers. In some cases, it is important to separate liquid and gas flowing simultaneously through a pipe. This is because the

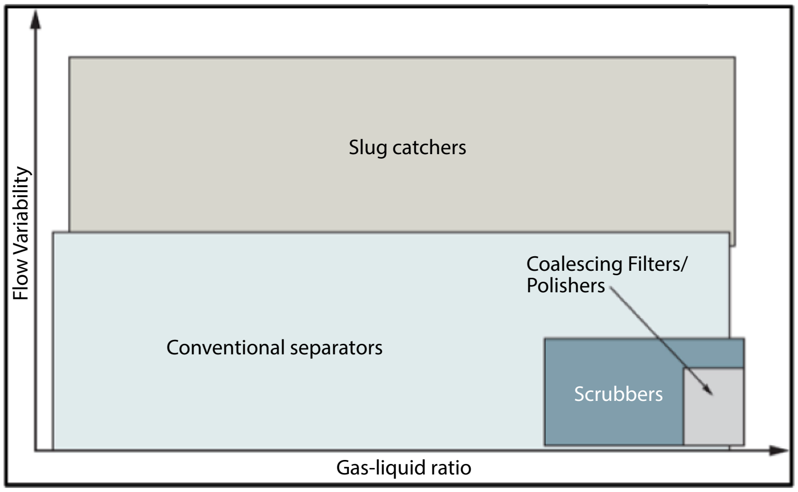


Figure 18.46 Gas–liquid separation equipment selection map [80].

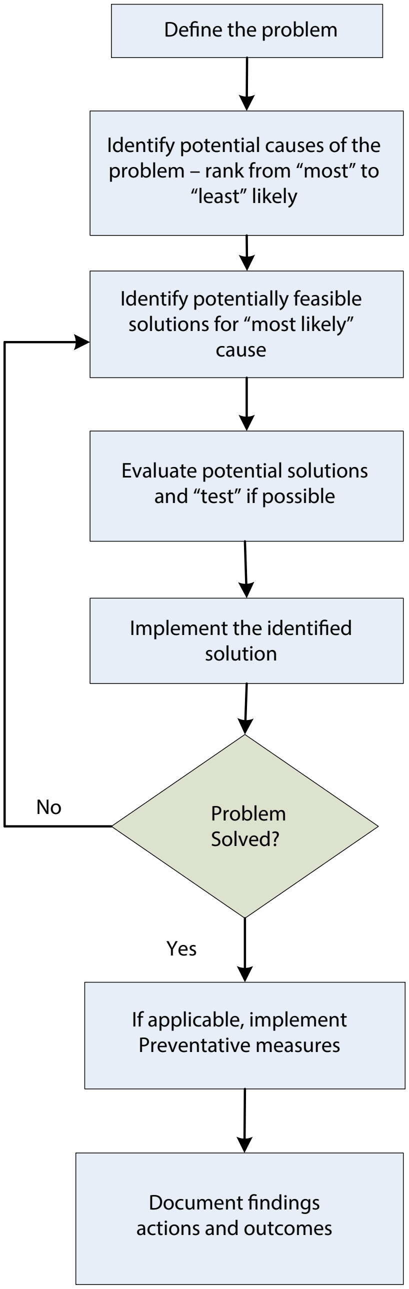


Figure 18.47 Troubleshooting methodology [80].

Table 18.14 Example of separator performance metrics [86].

	Value	
Parameter	**SI units**	**Imperial units**
Liquid content of separated gas	0.013–0.067 m^3 per 10^6 std m^3	0.1–0.5 gal per MMscf
Gas content of separated liquid	<1% volume/volume	<1% volume/volume
Target separable gas bubble size from liquid	200–300 μm	
Slug handling capacity	Application specific	

conditions of the flowing mixture and the efficiency of separation may vary widely. Therefore, a separator for such duty must be adequate. In addition, there are constraints due to space or weight that often affect the choice of separators, the need to handle solids or effect a three-phase separation, and the requirements for liquid hold-up. In practice, most separation problems are solved by the following types of separating equipment.

Knockout or Surge Drums (Figures 18.48 and 18.49). A knockout drum is suitable for a bulk separation of gas and liquid, particularly when the liquid volume fraction is high with stratified or plug flow in the pipe. Also, it is useful when vessel internals are required to be kept to a minimum, e.g., in relief systems or in fouling service. It is unsuitable if a mist is being separated or if high separating efficiency is required.

Cyclones. These are robust and not susceptible to fouling or wax (Figure 18.50). A multicyclone separator (Figure 18.51) is compact and is reasonably efficient for foam, but not for slugs. The efficiency of a cyclone decreases with increasing diameter, and cyclones are not applicable above 5 ft diameter. They possess a good efficiency for smaller flow rates. However, cyclones are expensive to operate, especially under vacuum conditions, because they have a higher pressure drop than either a knockout drum or a demister separator.

Demister Separators. A demister separator is fitted either with a vane demister package or with a wire demister mat. The latter type is much preferred, although it is unsuitable for fouling service. The wire mesh demister is a widely applied type of separator, and is adequate for all gas–liquid flow regimes over a wide range of gas flow rates. A knockout drum or demister separator may be either a vertical or a horizontal vessel. A vertical vessel is generally preferred because its efficiency does not vary with the liquid level. Alternatively, a horizontal vessel is chosen when it offers a clear size advantage, if the headroom is restricted or if a three-phase separation is required. Knockout drums and cyclones are recommended for waxy and coking feeds. Demister mats are not suitable because of the danger of plugging. Vane demister packages are used as alternatives, but provision should be made for cleaning.

Horizontal Versus Vertical Separators

Separators can be either horizontal or vertical. In the gravity section, droplets fall down against the direction of gas flow in the vertical separators whereas the droplets fall perpendicular to the direction of flow in horizontal separators. Whenever very high or very low gas-to-oil (GOR) is specified, including for scrubbers, vertical separators are preferred. For all other operations where separation is more difficult, horizontal separators are preferred. The following highlights the advantages of the horizontal over the vertical separations

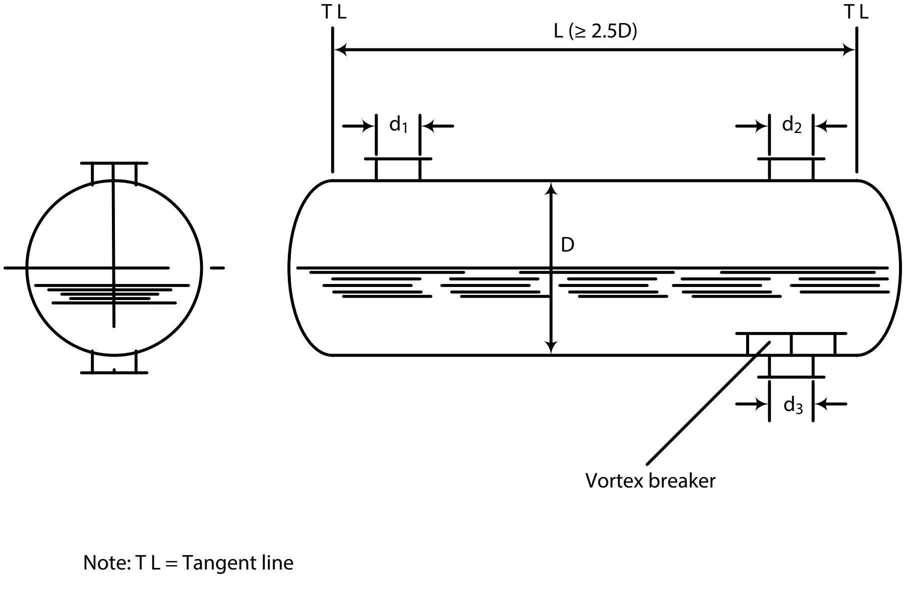


Figure 18.48 Horizontal knock-out drum.

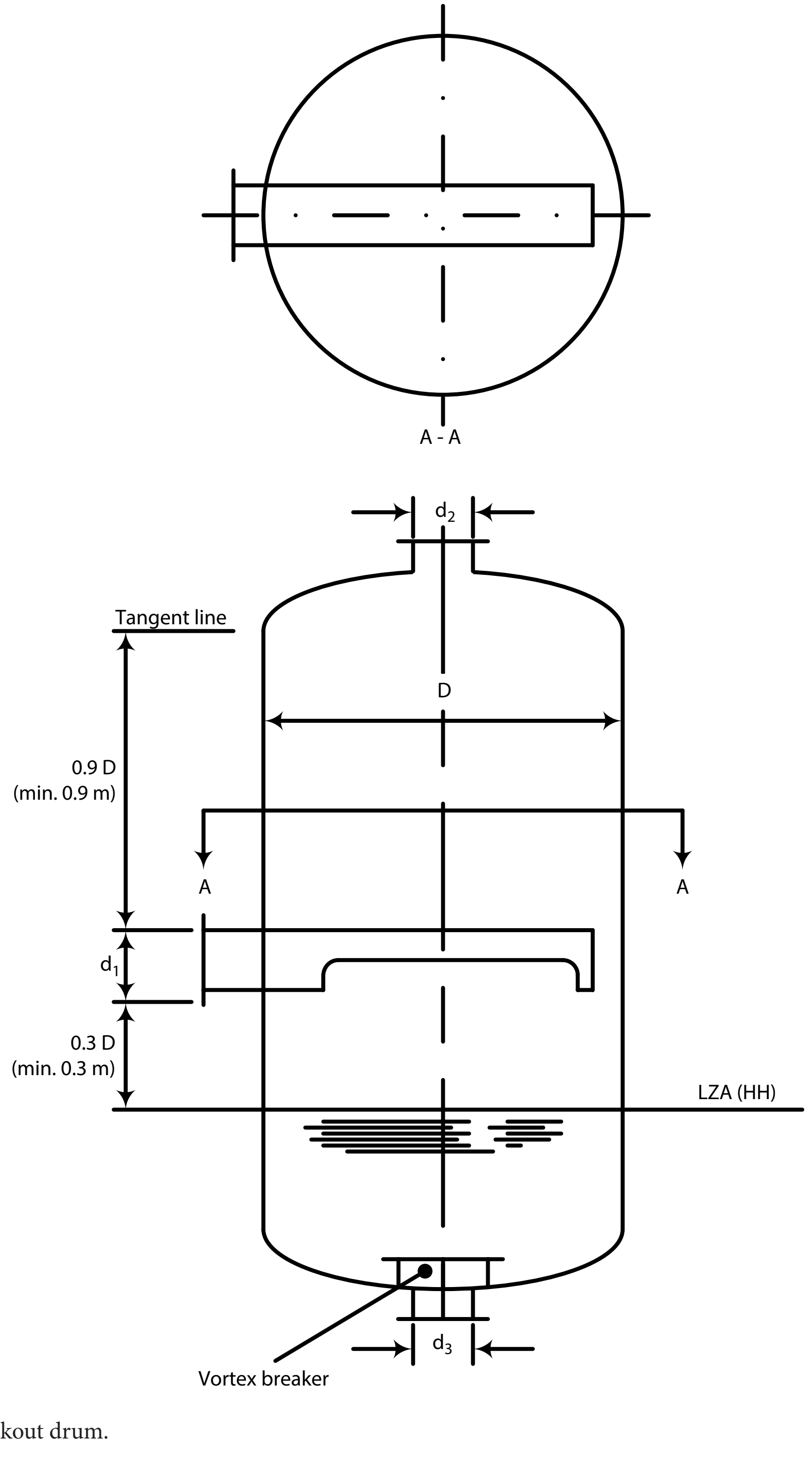


Figure 18.49 Vertical knockout drum.

- Liquid droplets are more efficiently separated from gas since they fall perpendicular to the flow of gas.
- Gas bubbles can break out of the liquid more easily because of the large gas–liquid interfacial area.
- Top platform of the separator is more easily accessible for operation/maintenance.

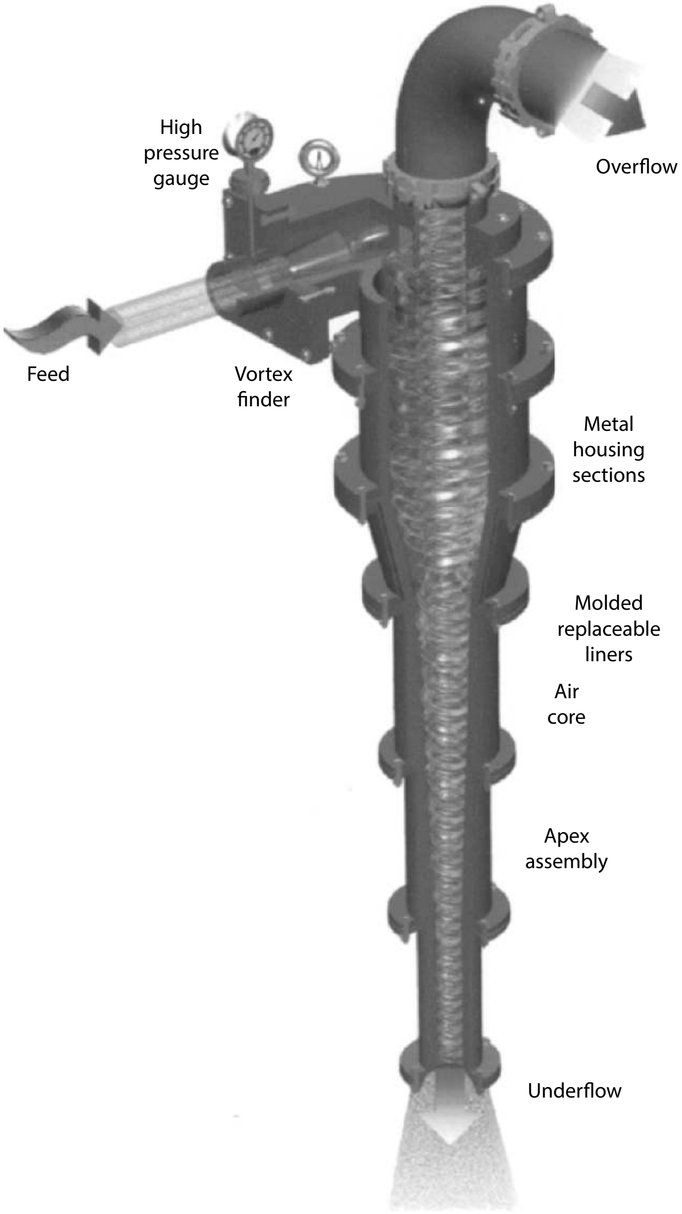


Figure 18.50 A cyclone separator (by permission from Krebs Engineers).

The horizontal separators have the following disadvantages over the vertical separators.

- The horizontal separators require more area than the vertical ones. Therefore, where area is critical, such as offshore locations, the vertical separators may be preferred.
- Sand separates easily in vertical separators since it falls to the bottom of the dish from where it can be removed. In horizontal separators, sand pans have to be installed with water flushing facility to remove the accumulated sand.
- Level control in horizontal vessels is more complicated than in vertical vessels.

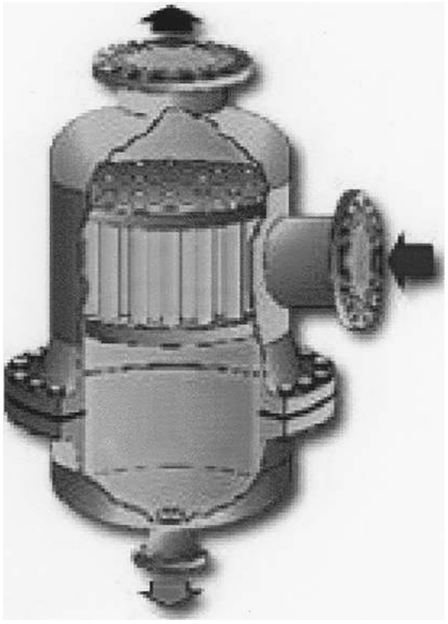


Figure 18.51 A multicyclone separator (by permission from Krebs Engineers).

Sizing of Vertical and Horizontal Separators

Vertical Separators. Vertical liquid–vapor separators are used to disengage a liquid from a vapor when the volume of liquid is small compared with that of vapor volume. A typical example of a vertical-liquid separator with small surge volume is a compressor knockout drum. The maximum allowable vapor velocity in a vertical separator that reduces the liquid carryover is dependent on the following:

- Liquid and vapor densities.
- A constant K based on surface tension, droplet size, and physical characteristics of the system.

The proportionality constant, K is 0.35 for oil and gas systems with at least 10 in. disengaging height between the mist eliminator bottom and gas liquid interface. For vertical vessels, K can vary between 0.1 and 0.35, if mist eliminators (demisters) are used to enhance disentrainment. The value of K also depends on the operating pressure of the vessel. At pressures above 30 psig, the K value decreases with pressure with an approximate value of 0.30 at 250 psig

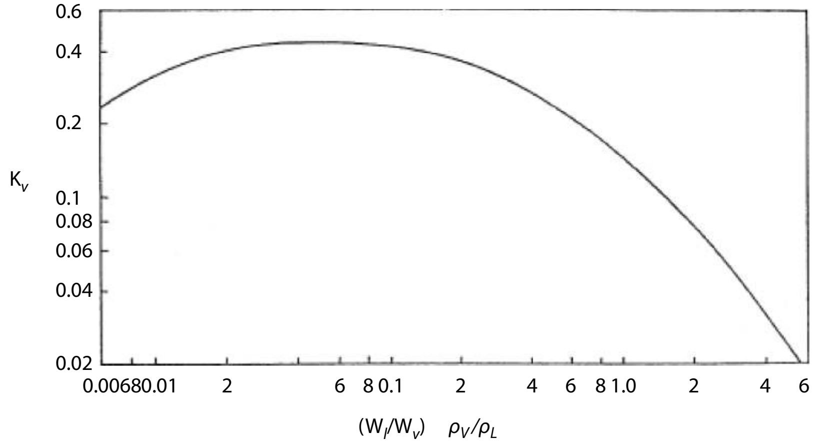


Figure 18.52 Design vapor velocity factor for vertical vapor–liquid separator at 85% of flooding (source: Watkins [34]).

and 0.275 at 800 psig. Watkins [34] has developed a correlation between the separation factor and K. Figure 18.52 illustrates Watkins's vapor velocity factor chart, based on five percent of the liquid being entrained with the vapor.

Branan [35] has developed a polynomial equation using Watkins's data to calculate the K value for a range of separating factors between 0.006 and 5.0. Watkins proposed a method for sizing reflux drums based upon several factors,

Table 18.15 Design criteria for reflux distillate accumulators.

			Minutes		
	Instrument factor F_1		**Labor factor F_2**		
Operation	**W/ alarm**	**W/o alarm**	**Good**	**Fair**	**Poor**
FRC	½	1	1	1.5	2
LRC	1	1½	1	1.5	2
TRC	1½	2	1	1.5	2

Table 18.16 Operation factors for external units.

Operating characteristics	**Factor F_3**
Under good control	2
Under fair control	3
Under poor control	4
Feed to or from storage	1.25
	Factor F_4
Board Mounted Level recorder	1.0
Level indicator on board	1.5
Gauge glass at equipment only	2.0

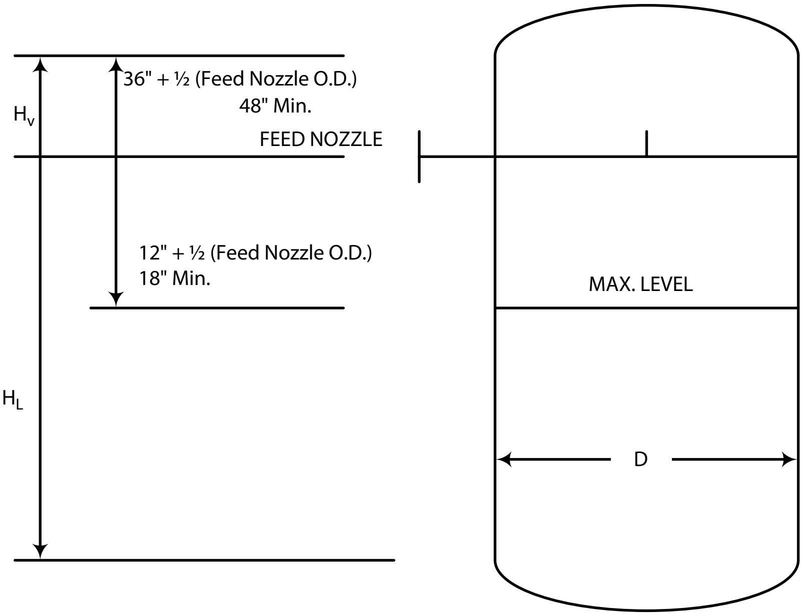


Figure 18.53 Dimensions of a vertical separator.

as illustrated in Tables 18.15 and 18.16. Table 18.15 gives the recommended design surge times. Table 18.16 gives the multiplying factors for various operator efficiencies. The operating factor is based upon the external unit and its operation, its instrumentation and response to control, the efficiency of labor, and chronic mechanic problems, and the possibility of short or long-term interruptions [34].

The multiplying factors F_1 and F_2 represent the instrument and the labor factors. A multiplying factor F_3, is applied to the net overhead product going downstream. F_4 depends on the kind and location of level indicators. It is recommended that 36 in. plus one-half the feed nozzle, OD, (48 in. minimum) be left above the feed nozzle for vapor. Below the feed nozzle, allowance of 12 in. plus one-half the feed nozzle, OD is required for clearance between the maximum liquid level and the feed nozzle (minimum of 18 in.). At some value between L/D ratios of 3 and 5, a minimum vessel weight will occur resulting in minimum costs for the separator. Figure 18.53 shows the dimensions of a vertical separator.

Calculation Method for a Vertical Drum

The following steps are used to size a vertical drum.

1. Calculate the vapor–liquid separation factor.

$$\text{S.Fac} = \frac{W_L}{W_V}\left(\frac{\rho_V}{\rho_L}\right)^{0.5} \tag{18.76}$$

2. From correlation, determine the design vapor velocity factor, K_v and the maximum design vapor velocity [35]

$$X = \ln(\text{S.Fac}) \tag{18.77}$$

$$K_v = \exp(A + B\,X + CX^2 + DX^3 + EX^4 + FX^5) \tag{18.78}$$

where

A = −1.942936
B = −0.814894
C = −0.179390
D = −0.0123790
E = 0.000386235
F = 0.000259550

$$V_{max} = K_v\left(\frac{\rho_L - \rho_V}{\rho_V}\right)^{0.5} \tag{18.79}$$

3. Calculate the minimum vessel cross sectional area.

$$Q_v = \frac{W_V}{3600(\rho_V)},\ \text{ft}^3/\text{s} \tag{18.80}$$

where Q_v = volumetric flow rate, ft^3/s.

$$A_v = \frac{Q_v}{V_{max}},\ \text{ft}^2 \tag{18.81}$$

4. Set a vessel diameter based on 6-inch increments and calculate cross sectional area.

$$D_{min} = \left(\frac{4.A_v}{\pi}\right)^{0.5}, \text{ft.} \tag{18.82}$$

$D = D_{min}$ to next largest 6 in.

$$\text{Area} = \frac{\pi D_{min}^2}{4} \tag{18.83}$$

5. Estimate the vapor–liquid inlet nozzle based on the following velocity criteria.

$$(U_{max})_{nozzle} = \frac{100}{\rho_{mix}^{0.5}}, \text{ft/s} \tag{18.84}$$

$$(U_{min})_{nozzle} = \frac{60}{\rho_{mix}^{0.5}}, \text{ft/s} \tag{18.85}$$

where

U_{max} = maximum velocity, ft/s

U_{min} = minimum velocity, ft/s

The density of the mixture ρ_{mix} is

$$\rho_{mix} = \left(\frac{W_L + W_V}{\frac{W_L}{\rho_L} + \frac{W_V}{\rho_V}}\right), \text{lb/ft}^3 \tag{18.86}$$

6. From Figure 18.53, make a preliminary vessel sizing for the height above the center line of a feed nozzle to top seam. Use 36 in. + ½ feed nozzle, OD or 48 in. minimum. Use 12 in. + ½ feed nozzle OD or 18 in. minimum to determine the distance below the center line of the feed nozzle to the maximum liquid level.
7. From Table 18.15 or 18.16, select the appropriate full surge volume in seconds. Calculate the required vessel volume as follows:

$$Q_L = \frac{W_L}{3600(\rho_L)}, \text{ft}^3/\text{s} \tag{18.87}$$

$$\begin{aligned} V &= (Q_L)(\text{design time to fill}), \text{ft}^3 \\ &= (60)(Q_L)(T), \text{ft}^3 \end{aligned} \tag{18.88}$$

The liquid height is:

$$H_L = V\left(\frac{4}{\pi D^2}\right), \text{ft} \tag{18.89}$$

8. Check geometry. $(H_L + H_V)/D$ must be between 3 and 5, where H_v is the vapor height in feet. For small volumes of liquid, it may be necessary to provide more liquid surge than is necessary to satisfy the L/D > 3. Otherwise this criterion should be observed. If the required liquid surge volume is greater than that obtained in a vessel having L/D < 5, a horizontal drum must be provided.

Example 18.6

Size a vertical separator under the following conditions

$$W_L = 5000\,\text{lb/h} \qquad W_V = 37{,}000\,\text{lb/h}$$

$$\rho_L = 61.87\,\text{lb/ft}^3 \qquad \rho_V = 0.374\,\text{lb/ft}^3$$

The liquid surge time is 5 min.

Solution

Calculate the vapor–liquid separation factor using Eq. 18.76.

$$\text{S.Fac} = \frac{5000}{37{,}000}\left(\frac{0.374}{61.87}\right)^{0.5} = 0.0105$$

From correlation, determine the design vapor velocity factor, K_v and the maximum design vapor velocity [35].

From Eq. 18.77

$$\begin{aligned} X &= \ln(0.0105) \\ &= -4.55574 \end{aligned}$$

From Eq. 18.78

$$\begin{aligned} K_v = \exp\,[&-1.942936 + (-0.814894)(-4.55575) + (-0.17939)(-4.55575)^2 \\ &+ (-0.012379)(-4.55575)^3 + (0.000386235)(-4.55575)^4 \\ &+ (0.00025955)(-4.55575)^5] \end{aligned}$$

$$\begin{aligned} K_v &= [-1.942936 + 3.712453 - 3.723214 + 1.170483 + 0.166376 - 0.509355] \\ K_v &= \exp[-1.126193] \\ &= 0.32426 \end{aligned}$$

From Eq. 18.79

$$V_{max} = 0.324\left(\frac{61.87 - 0.374}{0.374}\right)^{0.5} = 4.158\,\text{ft/s}$$

Calculate the minimum vessel cross sectional area using Eq. 18.80.

$$Q_v = \frac{37{,}000}{3600(0.374)} = 27.48\,\text{ft}^3/\text{s}$$

From Eq. 18.81,

$$A_v = \frac{27.48}{4.158} = 6.609\,\text{ft}^2$$

Set a vessel diameter based on 6-in. increments and calculate cross sectional area.

From Eq. 18.82

$$D_{\text{min}} = \left(\frac{4 \times 6.609}{\pi}\right)^{0.5} = 2.9\,\text{ft}$$

$D = D_{\text{min}}$ to next largest 6 in.

The density of the gas–liquid mixture ρ_{mix} from Eq. 18.86 is:

$$\rho_{\text{mix}} = \left(\frac{[5{,}000 + 37{,}000]}{\left[\frac{5{,}000}{61.87}\right] + \left[\frac{37{,}000}{0.374}\right]}\right),\ \text{lb}/\text{ft}^3 = 0.424\,\text{lb}/\text{ft}^3,\ \text{lb/ft}^3$$

Estimate the vapor–liquid inlet nozzle using Eqs. 18.84 and 18.85.

$$(U_{\text{max}})_{\text{nozzle}} = \frac{100}{(0.424)^{0.5}} = 153.57\ \text{ft/s}$$

$$(U_{\text{max}})_{\text{nozzle}} = \frac{60}{(0.424)^{0.5}} = 92.14\ \text{ft/s}$$

Calculate the required vessel volume using Eq. 18.87:

$$Q_L = \frac{5{,}000}{(3600 \times 61.87)} = 0.0224\,\text{ft}^3/\text{s}$$

From Eq. 18.88, the vessel volume V is:

$$V = (60)(Q_L)(T),\ \text{ft}^3$$
$$V = (60)\ (0.0224)\ (5)$$
$$= 6.73\ \text{ft}^3$$

The liquid height H_L is calculated using Eq. 18.89.

$$H_L = (6.72)\left(\frac{4}{\pi \times 2.9^2}\right) = 1.017\,\text{ft}$$

The Excel spreadsheet Example 18.6.xlsx has been developed for sizing gas–liquid vertical separator. The results of the Excel spreadsheet calculations are shown below.

RESULTS OF VERTICAL SEPARATOR SIZING		
LIQUID FLOW RATE	5000	lb/h
VAPOR FLOW RATE	37000	lb/h
LIQUID DENSITY	61.87	lb/ft^3
VAPOR DENSITY	0.374	lb/ft^3
LIQUID VOLUMETRIC RATE	0.022	ft^3/s
VAPOR VOLUMETRIC RATE	27.48	ft^3/s
LIQUID SURGE TIME	5	min.
SEPARATION FACTOR	0.011	
VAPOR VELOCITY FACTOR	0.331	
MAXIMUM VAPOR VELOCITY	4.24	ft/s
VESSEL DIAMETER	2.8726	ft.
MAXIMUM VAPOR-LIQUID NOZZLE VELOCITY	153.54	ft/s
MINIMUM VAPOR-LIQUID NOZZLE VELOCITY	92.12	ft/s
REQUIRED VESSEL VOLUME	6.6	ft^3
LIQUID HEIGHT	1.02	ft

Calculation Method for a Horizontal Drum

Horizontal vessels are used for substantial vapor–liquid separation where the liquid holdup space must be large. Maximum vapor velocity and minimum vapor space are determined as in the vertical drum, except that K_H for horizontal separators is generally set at 1.25 K_v. The following steps are carried out in sizing horizontal separators.

1. Calculate the vapor–liquid separation factor by Eq. 18.76 and K_v by Eqs. 18.77 and 18.78.
2. For horizontal vessels

$$K_H = 1.25K_v \tag{18.90}$$

3. Calculate the maximum design vapor velocity.

$$(U_V)_{max} = K_H\left(\frac{\rho_L - \rho_V}{\rho_V}\right)^{0.5}, \text{ft/s} \tag{18.91}$$

4. Calculate the required vapor flow area.

$$(A_V)_{min} = \frac{Q_v}{(U_V)_{max}}, \text{ft}^2 \tag{18.92}$$

5. Select appropriate design surge times from Table 18.15 or Table 18.16 and calculate full liquid volume by Eqs. 18.87 and 18.88. The remainder of the sizing procedure is carried out by trial and error as follows:
6. When the vessel is full, the separator vapor area can be assumed to occupy only 15 to 25% of the total cross-sectional area. Here, a value of 20% is used and the total cross-sectional area is expressed as:

$$(A_{Total})_{min} = \frac{(A_v)_{min}}{0.2} \tag{18.93}$$

$$D_{min} = \left(\frac{4(A_{total})_{min}}{\pi}\right)^{0.5} \tag{18.94}$$

7. Assume the length-to diameter ratio of 3 (i.e., L/D = 3). Calculate the vessel length.

$$L = (3)(D_{min}) \tag{18.95}$$

8. Because the vapor area is assumed to occupy 20% of the total cross-sectional area, the liquid area will occupy 80% of the total cross-sectional area.

$$A_L = (0.8)(A_{total})_{min}, \text{ ft}^2 \tag{18.96}$$

9. Calculate the vessel volume.

$$V_{VES} = (A_{total})_{min}(L), \text{ ft}^3 \tag{18.97}$$

10. Calculate liquid surge time.

$$T = \frac{(60)(A_L)(L)(\rho_L)}{W_L}, \text{ min.} \tag{18.98}$$

11. If $5 < L / D < 3$, resize.

If there is water to be settled and withdrawn from hydrocarbon, the water's settling time requirement needs to be checked. The water settling requirement rather than other process considerations might set the liquid surge capacity. Therefore, the liquid surge capacity that has been previously estimated from tables might have to be increased.

The following is a quick check for water settling [35]:

1. Estimate the water terminal settling velocity using

$$V_T = 44.7 \times 10^{-8} \frac{(\rho_w - \rho_o)F_s}{\mu_o} \tag{18.99}$$

where

V_T = terminal settling velocity, ft/s
F_s = correction factor for hindered settling
ρ_w = density of water, lb/ft^3
ρ_o = density of oil, lb/ft^3

μ_o = absolute viscosity of oil, lb/ft.-s

This assumes a droplet diameter of 0.0005 ft (152 μm). F_s is determined from:

$$F_s = \frac{X^2}{10^{1.82(1-X)}} \tag{18.100}$$

where X = volume fraction of oil.

2. Calculate the modified Reynolds number, Re from:

$$Re = 5 \times 10^{-4} \frac{\rho_o V_T}{\mu_o} (\text{usually} < 1.0) \tag{18.101}$$

This assumes a droplet diameter of 0.0005 ft (152 μm)

3. Calculate V_s/V_T from:

$$V_s/V_T = A + B(\ln Re) + C(\ln Re^2) + D(\ln Re^3) + E(\ln Re^4) \tag{18.102}$$

where

V_s = actual settling velocity, ft/s
A = 0.919832
B = −0.091353
C = −0.017157
D = 0.00029258
E = −0.00011591

4. Calculate the length of the settling section (Figure 18.54a) as:

$$L = \frac{hQ}{AV_s} \tag{18.103}$$

where

L = length of settling zone, ft.
h = height of oil, ft.
Q = flow rate, ft^3/s

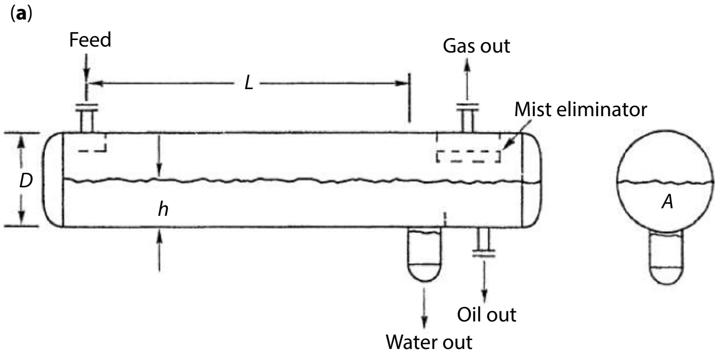


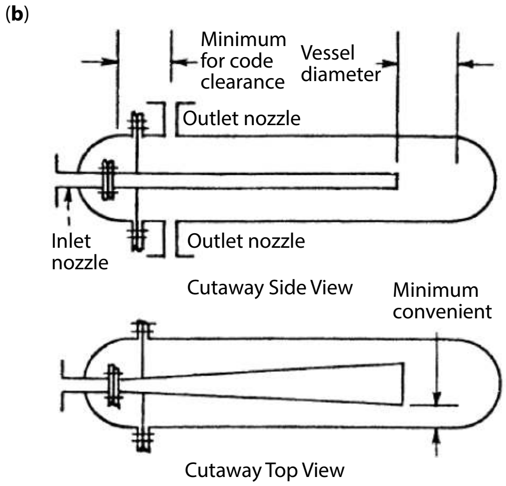
Figure 18.54a Typical vapor–liquid separator [35].

A = cross sectional area of the oil settling zone, ft^2

This allows the water to fall out and be drawn off at the bootleg before leaving the settling section.

Liquid–Liquid Separation

The separation of immiscible liquids is an essential chemical/process engineering operation. Barton [36] provides the following empirical method, which has proven satisfactory for sizing liquid–liquid phase separators.



Note:

1. For large vessels, a manway and internally dismantled cone may be used where more economical.
2. Inlet and outlet nozzles should be sized for pump discharge.
3. Gauge glass and level instruments should be located at inlet-outlet end.
4. Mechanical design should suit economy under operating conditions.

Figure 18.54b Recommended design for a liquid–liquid separator [36].

Figure 18.55 A photograph of a horizontal separator in a refinery facility.

Figure 18.56 A photograph of a vertical separator (source: www.spe.org).

1. Calculate the holdup time as:

$$T = 0.1\left[\frac{\mu}{(SpGr_b - SpGr_t)}\right] \tag{18.104}$$

where

T = holdup time, h.
μ = viscosity of the continuous phase, cP
$SpGr_b$ = specific gravity of the bottom phase
$SpGr_t$ = specific gravity of the top phase

2. Assign a length-to-diameter ratio of 5, and size a tank to accommodate the required holdup time.
3. Provide inlet and outlet nozzles at one end, and an internal flat cone (see Figure 18.54b).

The relation between viscosity and specific gravity difference of the phases corresponds to those of the equations for terminal settling velocity in the Stokes' law region and free-settling velocity of isomeric particles. Also, the dimensions of the tank and cone recognize that the shape of turbulence created by nozzles discharging into liquids spreads at an angle whose slope is about 1–5. This method is not applicable for emulsions.

Figure 18.55 shows a photograph of a horizontal separator in a refinery facility, and Figure 18.56 shows a photograph of a vertical separator.

Example 18.7

Size a horizontal separator for the following conditions

$$W_L = 56{,}150\,\text{lb/h} \qquad W_V = 40{,}000\,\text{lb/h}$$

$$\rho_L = 60.0\,\text{lb/ft}^3 \qquad \rho_V = 1.47\,\text{lb/ft}^3$$

The liquid surge time is 6 min. Use L/D = 3.0.

Solution

Calculate the vapor–liquid separation factor by Eq. 18.76 and K_v by Eqs. 18.77 and 18.78.

$$= \text{S.Fac} = \frac{56{,}150}{40{,}000}\left(\frac{1.47}{60.0}\right)^{0.5} = 0.21972$$

From correlation, determine the design vapor velocity factor, K_v and the maximum design vapor velocity [35]

From Eq. 18.77

$$X = \ln(0.21972) = -1.51539$$

From Eq. (18.78)

$$\begin{aligned} K_v = \exp\,[&-1.942936 + (-0.814894)(-1.51539) + (-0.17939)(-1.51539)^2 \\ &+ (-0.012379)(-1.51539)^3 + (0.000386235)(-1.51539)^4 \\ &+ (0.00025955)(-1.51539)^5] \end{aligned}$$

$$K_v = [-1.942936 + 1.234882 - 0.411952 + 0.043078 + 0.002037 - 0.000207]$$

$$K_v = \exp[-1.075098] = 0.34126$$

From Eq. 18.90, for horizontal vessels, K_H is:

$$K_H = 1.25(0.341) = 0.426$$

Calculate the maximum design vapor velocity.

From Eq. 18.91

$$(U_V)_{max} = 0.426\left(\frac{60-1.47}{1.47}\right)^{0.5} = 2.69\ \text{ft/s}$$

Calculate the required vapor flow area using Eq. 18.92.

$$Q_v = \frac{W_V}{(3600)(\rho_v)} = \frac{40{,}000}{(3600)(1.47)} = 7.56\,\text{ft}^3/\text{s}$$

$$(A_v)_{min} = \frac{7.56}{2.69} = 2.81\,\text{ft}^2$$

When the vessel is full, the separator vapor area can be assumed to occupy only 15–25% of the total cross-sectional area. Here, a value of 20% is used and the total cross-sectional area is expressed as follows (using Eqs. 18.93 and 18.94).

$$(A_{Total})_{min} = \frac{2.81}{0.2} = 14.05\,\text{ft}^2$$

$$D_{min} = \left[\frac{4 \times 14.05}{\pi}\right]^{0.5} = 4.23\,\text{ft}$$

Assume the length-to diameter ratio of 3 (i.e., L/D = 3). Calculate the vessel length.

From Eq. 18.95,

$$L = (3)\ (4.23) = 12.7\ \text{ft}.$$

Because the vapor area is assumed to occupy 20% of the total cross-sectional area, the liquid area will occupy 80% of the total cross-sectional area.

From Eq. 18.96

$$A_L = (0.8)\ (14.05) = 11.24\ \text{ft}^2$$

Calculate the vessel volume from Eq. 18.97.

$$V_{VES} = (14.05)\ (12.7) = 178.435\ \text{ft}^3$$

Calculate liquid surge time using Eq. 18.98.

$$T = \frac{(60)(11.24)(12.7)(60)}{56150} = 9.15\,\text{min}.$$

The Excel spreadsheet Example 18.7.xlsx has been developed to size horizontal gas-liquid separator. The results of the Excel spreadsheet calculations are shown below.

RESULTS OF HORIZONTAL SEPARATOR SIZING		
LIQUID FLOW RATE	56150	lb/h
VAPOR FLOW RATE	40000	lb/h
LIQUID DENSITY	60	lb/ft^3
VAPOR DENSITY	1.47	lb/ft^3
LIQUID VOLUMETRIC RATE	0.26	ft^3/s
VAPOR VOLUMETRIC RATE	7.56	ft^3/s
LIQUID SURGE TIME	6	min.
LENGTH TO DIAMETER RATIO (L/D)	3	
SEPARATION FACTOR	0.22	
VAPOR VELOCITY FACTOR	0.426	
VAPOR VOLUMETRIC RATE	7.56	ft^3/s
MAXIMUM VAPOR VELOCITY	2.69	ft/s
VESSEL DIAMETER	4.23	ft

VESSEL LENGTH	12.69	ft
VESSEL VOLUME	178.32	ft^3
CALCULATED LIQUID SURGE TIME	9.15	min.

Example 18.8 Horizontal Gravity Separator Without Mist Extractor (GPSA)

A horizontal gravity separator (without mist extractor) is required to handle 60 MMscfd of 0.75 specific gravity gas (MW = 21.72) at a pressure of 500 psig and a temperature of 100°F. Compressibility factor is 0.9, viscosity is 0.012cP, and liquid specific gravity is 0.5. It is desired to remove all entrainment greater than 150 μm in diameter. No liquid surge is required.

Solution

$$\text{Gas density}, \rho_g = \frac{P(M_w)}{ZRT} = \frac{(514.7)(21.72)}{(0.90)(10.73)(560)} = 2.07\,\text{lb}/\text{ft}^3$$

$$\text{Liquid density}, \rho_l = (0.5)(62.4) = 31.2\,\text{lb}/\text{ft}^3$$

$$\text{Mass flow}, G = \frac{(60)(10^6)(21.72)}{(379)(24)(3600)} = 39.8\,\text{lb}/\text{s}$$

$$\text{Particle diameter}, D_p = \frac{(150)(0.00003937)}{12} = 0.000492\,\text{ft}$$

From Eq. 18.27

$$C'(\text{Re})^2 = \frac{(0.95)(10^8)\rho_g D_p^3(\rho_l - \rho_g)}{\mu^2}$$

$$= \frac{(0.95)(10^8)(2.07)(0.000492)^2(31.2 - 2.07)}{(0.012)^2} = 4738$$

From Figure 18.8, Drag coefficient C′ = 1.40

$$\text{Terminal velocity}, V_t = \sqrt{\frac{4g D_p(\rho_l - \rho_g)}{3\rho_g C'}}$$

$$= \sqrt{\frac{4(32.2)(0.000492)(29.13)}{3(2.07)(1.40)}} = 0.46\,\text{ft/s}$$

$$\text{Gas flow rate}, Q_G = \frac{G}{\rho_G} = \frac{39.80}{2.07} = 19.2\,\text{ft}^3/\text{s}$$

Assume a diameter, D_v = 3.5 ft

$$\text{Vessel length}, L = \frac{4Q_G}{\pi V_t D_v} = \frac{4(19.2)}{\pi(0.46)(3.5)} = 15.2\,\text{ft}$$

Other reasonable solutions are as follows:

Diameter, ft	Length, ft
3.5	15.2
4.0	13.3
4.5	11.8
5.0	10.6
5.5	9.6
6.0	8.9

Example 18.9

What size vertical separator without a mist extractor is required to meet the conditions used in Example 18.8?

Solution

$$\text{Area, A} = \frac{Q_G}{V_t} = \frac{19.2}{0.46} = 41.7\,\text{ft}^2$$

$$\text{Diameter, } D_v = \sqrt{\frac{4\,A}{\pi}} = \sqrt{\frac{4(41.7)}{\pi}} = 7.29\text{ ft minimum}$$

Use a 90-in. ID vertical separator.

Liquid Holdup and Vapor Space Disengagement

The dimensions of both vertical and horizontal separators are based on rules designed to provide adequate liquid holdup and vapor disengaging space. For instance, the desired vapor space in a vertical separator is at least 1 ½ times the diameter, with 6 in. as the minimum above the top of the inlet nozzle. In addition, a 6 in. minimum is required between the maximum liquid level and the bottom of the inlet nozzle. For a horizontal separator, the minimum vapor space is equal to 20% of the diameter of 12 in.

Wire Mesh Pad

Pads of fine wire mesh indicate coalescence of impinging droplets into larger ones, which then separate freely from the gas phase. No standard equations have been developed for the pressure drop across wire mesh because there are no standardized mesh pads. However, as a rule of thumb, the pressure drop of a wire mesh is $\Delta P = 1.0$ in. H_2O. Every manufacturer makes a standard high efficiency, very high efficiency, or high throughput mesh under various trade names, each for a specific requirement.

Standards for Horizontal Separators

The following specifications are generally standard in the design of horizontal separators [37].

1. The maximum liquid level shall provide a minimum vapor space height of 15 in. (381 mm) but should not be below the center line of the separator.

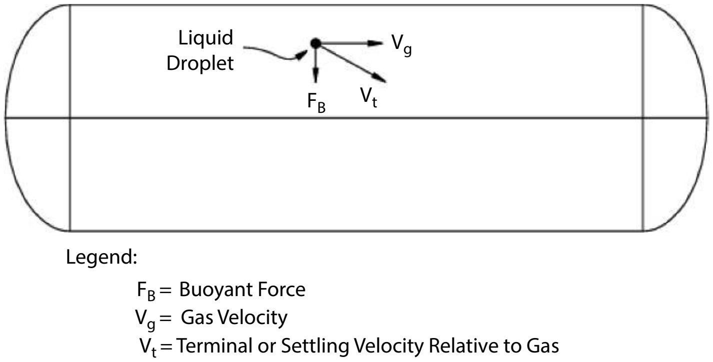


Figure 18.57 Model for a horizontal separator [79].

2. The volume of dished heads is not considered in vessel sizing calculations.
3. The inlet and outlet nozzles shall be located as closely as practical to the vessel tangent lines. Liquid outlets shall have anti-vortex baffles.

Sizing Horizontal Separators

In sizing a horizontal separator, it is essential to choose a seam-to-seam vessel length and a diameter. This is to satisfy the conditions for gas capacity that allows the liquid droplets to fall from the gas to the liquid volume as the gas traverses the effective length of the vessel. It should also provide adequate retention time to allow the liquid to reach equilibrium. Figure 18.57 shows a vessel 50% full of liquid, which is used to develop sizing equations for a horizontal separator.

Gas Capacity Constraint

The principles of liquid droplets settling through a gas can be used to develop an equation to size a separator for a gas flow rate. The gas capacity equations are based on setting the gas retention time equal to the time required for a droplet to settle to the liquid interface. The equations are derived for a vessel 50% full of liquid and separation of 100-μm liquid droplets from the gas [79].

Field Units

$$dL_{eff} = 420\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \qquad (18.105)$$

SI Units

$$dL_{eff} = 34.5\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \qquad (18.106)$$

where

d = vessel internal diameter, in. (mm)
L_{eff} = effective length of the vessel where separation occurs, ft (m)
T = operating temperature, °R (K)

P = operating pressure, psia (kPa)
Z = gas compressibility
Q_g = gas flow rate, MMscfd (scm/h)
C_D = drag coefficient
d_m = liquid droplet to be separated, (μm)
ρ_g = density of gas, lb/ft³ (kg/m³)
ρ_1 = density of liquid, lb/ft³ (kg/m³)

Assuming that the vessel is half full of liquid

where

Area, A = ft² (m²)
Diameter, D = ft. (m)
Diameter, d = in. (mm)
Volumetric flow rate, Q = ft³/s (m³/s)

Field Units

Determine the gas velocity, V_g:

$$V_g = \frac{Q}{A_g} \tag{18.107}$$

$$\begin{aligned} A_g &= \frac{1}{2}\left(\frac{\pi D^2}{4}\right) \\ &= \frac{1}{2}\left(\frac{\pi}{4}\frac{d^2}{144}\right) = \frac{d^2}{367} \end{aligned} \tag{18.108}$$

Q_g is in MMscfd

$$\begin{aligned} Q &= Q_g \times 10^6 \left(\frac{\text{scf}}{\text{MMscf}}\right)\left(\frac{\text{day}}{24\text{h}}\right)\left(\frac{\text{h}}{3600\text{s}}\right) \times \frac{14.7}{P} \times \frac{TZ}{520} \\ &= 0.327\frac{TZ}{P}Q_g \end{aligned} \tag{18.109}$$

$$V_g = \frac{\left(0.327\frac{TZ}{P}Q_g\right)(367)}{d^2} = 120\frac{TZQ_g}{Pd^2} \tag{18.110}$$

SI Units

From Eq. 18.107:

$$V_g = \frac{Q}{A_g}$$

$$A_g = \frac{1}{2}\left(\frac{\pi D^2}{4}\right)$$
$$= \frac{1}{2}\left(\frac{\pi}{4}\frac{d^2}{1000^2}\right) = 3.927 \times 10^{-7}\, d^2 \tag{18.111}$$

$$Q_g = \text{scm/h}$$
$$Q = Q_g \times \frac{101.3}{P} \times \frac{TZ}{288.6}\left(\frac{h}{3600s}\right) = 9.75 \times 10^{-5}\,\frac{TZ}{P}Q_g \tag{18.112}$$

$$V_g = \frac{\left(9.75 \times 10^{-5}\,\frac{TZ}{P}Q_g\right)}{3.927 \times 10^{-7}\, d^2} = 248.3\frac{TZQ_g}{Pd^2} \tag{18.113}$$

The residence time of the gas (t_g) is equal to the time required for the droplet (t_d) to fall to the gas–liquid interface:

Field Units

$$t_g = \frac{L_{eff}}{V_g} \qquad t_d = \frac{D}{2V_t} = \frac{d}{24V_t} \tag{18.114}$$

Substituting Eq. 18.110 into Eq. 18.144 gives

$$t_g = \frac{L_{eff}}{120\left(\frac{TZQ_g}{Pd^2}\right)} \tag{18.115}$$

The terminal velocity of gas, V_t is given by:

$$V_t = 0.0119\left[\left(\frac{\rho_l - \rho_g}{\rho^g}\right)\frac{d_m}{C_D}\right]^{0.5} \tag{18.116}$$

Substituting Eq. 18.116 into Eq. 18.144 gives:

$$t_d = \frac{d}{(24)(0.0119)}\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \tag{18.117}$$

Setting $t_g = t_d$ gives

$$\frac{L_{eff}}{120\left(\dfrac{TZQ_g}{Pd^2}\right)} = \frac{d\left[\left(\dfrac{\rho_g}{\rho_l - \rho_g}\right)\dfrac{C_D}{d_m}\right]^{0.5}}{(24)(0.0119)} \qquad (18.118)$$

Rearranging Eq. 18.118 gives:

$$L_{eff}d = 420\frac{TZQ_g}{P}\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \qquad (18.105)$$

SI Units

$$t_g = \frac{L_{eff}}{V_g} \qquad t_d = \frac{D}{2V_t} = \frac{d}{2000V_t} \qquad (18.119)$$

Substituting Eq. 18.113 into Eq. 18.119 gives

$$t_g = \frac{L_{eff}}{248.3\left(\dfrac{TZQ_g}{Pd^2}\right)} \qquad (18.120)$$

The terminal velocity of gas, V_t is given by:

$$V_t = 0.0036\left[\left(\frac{\rho_l - \rho_g}{\rho_g}\right)\frac{d_m}{C_D}\right]^{0.5} \qquad (18.121)$$

Substituting Eq. 18.121 into Eq. 18.119 gives:

$$t_d = \frac{d}{(2000)(0.0036)}\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \qquad (18.122)$$

Setting $t_g = t_d$ gives

$$\frac{L_{eff}}{248.3\left(\dfrac{TZQ_g}{Pd^2}\right)} = \frac{d\left[\left(\dfrac{\rho_g}{\rho_l - \rho_g}\right)\dfrac{C_D}{d_m}\right]^{0.5}}{(2000)(0.0036)} \qquad (18.123)$$

Rearranging Eq. 18.123 gives:

$$L_{eff}d = 34.5\frac{TZQ_g}{P}\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \tag{18.106}$$

Liquid Capacity Constraint

Two-phase separators must be sized to provide some liquid retention time so the liquid can reach phase equilibrium with the gas. For a vessel 50% full of liquid, with a specified liquid flow rate and retention time, the following may be used to determine a vessel size.

Field Units

$$d^2L_{eff} = \frac{t_r\,Q_l}{0.7} \tag{18.124}$$

SI Units

$$d^2Leff = 42{,}445\ t_r\ Q_l \tag{18.125}$$

where

t_r = desired retention time for the liquid, min.
Q_l = liquid flow rate, bpd (m^3/h)

Eqs. 18.124 and 18.125 are derived as follows:

where

time, t = s
Volume, V = ft^3 (m^3)
Volumetric flow rate, Q = ft^3/s ($m^3/min.$)

Field Units

$$t = \frac{V}{Q} \tag{18.126}$$

$$V = \frac{1}{2}\left(\frac{\pi D^2 L_{eff}}{4}\right) = \frac{\pi d^2 L_{eff}}{(2)(4)(144)} = 2.73\times10^{-3}\, d^2 L_{eff} \tag{18.127}$$

Q_l is in bpd

$$Q = Q_l \times 5.61\left(\frac{ft^3}{barrel}\right)\left(\frac{day}{24h}\right)\left(\frac{h}{3600s}\right) = 6.50 \times 10^{-5} Q_l \tag{18.128}$$

Substituting Eqs. 18.127 and 18.128 into Eq. 18.126 gives:

$$t = 42.0\frac{d^2 L_{eff}}{Q_l} \tag{18.129}$$

t_r is in min. = $60t_r$

$$d^2 L_{eff} = \frac{t_r Q_l}{0.7} \tag{18.130}$$

SI Units

$$V = \frac{1}{2}\left(\frac{\pi D^2 L_{eff}}{4}\right) = \frac{\pi L_{eff}}{8}\left(\frac{d}{1000}\right)^2 = 3.927 \times 10^{-7} d^2 L_{eff} \tag{18.131}$$

$$Q_l = m^3/min$$

$$Q = Q_l \times \left(\frac{h}{60\,min}\right) = \frac{Q_l}{60} \tag{18.132}$$

Substituting Eqs. 18.131 and 18.132 into Eq. 18.126 gives:

$$t = \frac{V}{Q} = \frac{3.9269 \times 10^{-7} d^2 L_{eff}}{\dfrac{Q_l}{60}} = 2.35619 \times 10^{-5}\frac{d^2 L_{eff}}{Q_l} \tag{18.133}$$

Rearranging Eq. 18.133 gives

$$d^2L_{eff} = 42{,}445\ t_r\ Q_l \tag{18.134}$$

Seam-to-Seam Length

The effective length may be calculated from Eqs. 18.105 and 18.106 and 18.124 and 18.125 from which a vessel seam-to-seam length can be determined. The actual required seam-to-seam length is dependent on the physical design of the internals of the vessel. For vessels sized on a gas capacity basis, some portion of the vessel length is required to distribute the flow evenly near the inlet diverter. The length of the vessel between the inlet diverter and the mist extractor with evenly distributed flow is the L_{eff} calculated from Eqs. 18.105 and 18.106. As a vessel's diameter increases, more length is required to evenly distribute the gas flow (Figure 18.58).

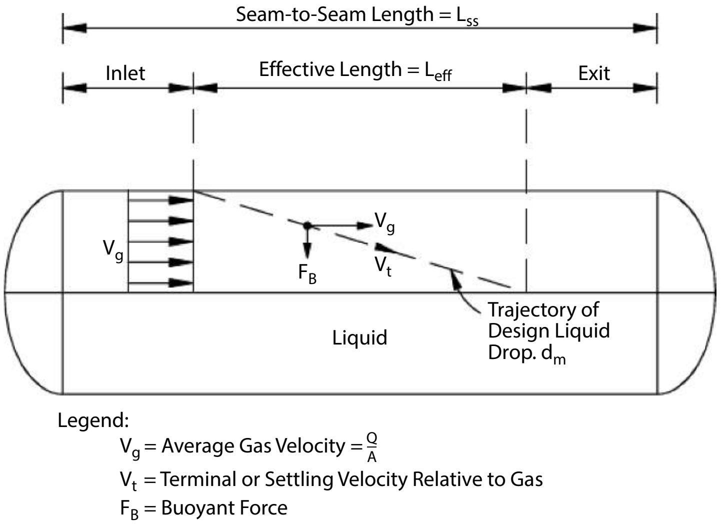


Figure 18.58 Approximate seam-to-seam length of a horizontal separator one-half full [79].

The seam-to-seam length L_{ss} is determined by:

Field Units

$$L_{ss} = L_{eff} + \frac{d}{12} \quad \text{for gas capacity} \tag{18.135}$$

SI Units

$$L_{ss} = L_{eff} + \frac{d}{1000} \quad \text{for gas capacity} \tag{18.136}$$

For vessels sized on a liquid capacity basis, some portion of the vessel length is required for inlet diverter flow distribution and liquid outlet. The seam-to-seam length should not exceed the following:

$$L_{ss} = \frac{4}{3} L_{eff} \tag{18.137}$$

Slenderness Ratio

Eqs. 18.105 and 18.106 and 18.124 and 18.125 allow for various choices of diameter and length. For each vessel design, a combination of L_{eff} and d exists that will minimize the cost of the vessel. The smaller the diameter, the less the vessel will weigh and thus the lower its cost. There is point where decreasing the diameter increases the possibility that high velocity in the gas flow will create waves and re-entrain liquids at the gas-liquid interface. Where the gas capacity governs and length divided by the diameter referred to as the "slenderness ratio" is greater than 4 or 5, re-entrainment could become a problem. Eq. 18.138 indicates that slenderness ratio must be at least 1 or more, and two-phase separators are designed for slenderness ratios between 3 and 4.

Field Units

$$S_R = \frac{12 L_{ss}}{d} \tag{18.138}$$

SI Units

$$S_R = \frac{1000 L_{ss}}{d} \tag{18.139}$$

Procedure for Sizing Horizontal Separators—Half Full [79]

The following steps are:

1. Establish the design basis. This includes specifying the maximum and minimum flow rates, operating temperature and pressure, droplet size to be removed, and so on.
2. Calculate values of L_{eff} for selected values of d that satisfy Eqs. 18.105 and 18.106 and the gas capacity constraint. Use Eqs. 18.135 and 18.136 to determine L_{ss}.
3. Calculate values of L_{eff} using Eqs. 18.124 and 18.125 for liquid capacity.
4. For each d, the larger value of L_{ss} should be used.
5. Calculate the slenderness ratio and select a combination of d and L_{ss} that has a slenderness ratio between 3 and 4. Lower ratios can be chosen if dictated by available space and higher ratios can be chosen if the vessel is checked for re-entrainment.
6. Make a final selection; always select a standard vessel size. Vessels with outside diameters up through 24 in. (600 mm) have nominal pipe dimensions. Vessels with outside diameters larger than 24 in. (600 mm) are rolled from plate with diameter increments of 6 in. (150 mm). The shell seam-to-seam length is expanded in 2.5 ft. (750 mm) segments and is usually from 5 to 10 ft (1500 to 3000 mm). Standard separator vessel sizes may be obtained from API 121J [75]. In larger sizes in most locations, heads come in outside diameters, which are multiples of 6 in. (150 mm). The width of steel sheets for the shells is usually 10 ft (3000 mm), thus it is common practice to specify L_{ss} in multiples of 5.

Horizontal Separators Sizing Other Than Half Full

For vessels that are designed other than 50% full, the following equations are used for the actual gas and liquid areas to calculate gas velocity and liquid volume (Figure 18.59):

Field units

$$L_{eff} d = 420 \left(\frac{1-\beta}{1-\alpha} \right) \left[\frac{TZQ_g}{P} \right] \left[\left(\frac{\rho_g}{\rho_l - \rho_g} \right) \frac{C_D}{d_m} \right]^{0.5} \tag{18.140}$$

where

$\left(\frac{1-\beta}{1-\alpha} \right)$ = design constant (Figure 18.60a)

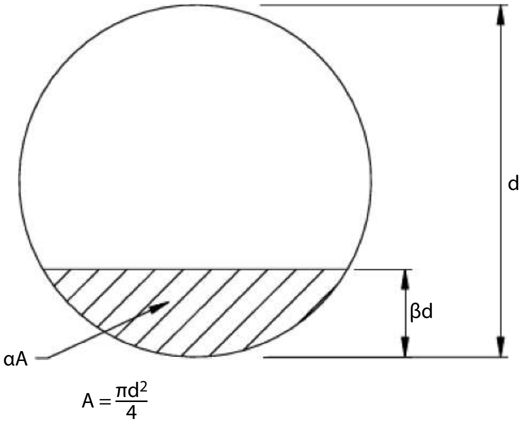


Figure 18.59 Definition of parallel areas [79].

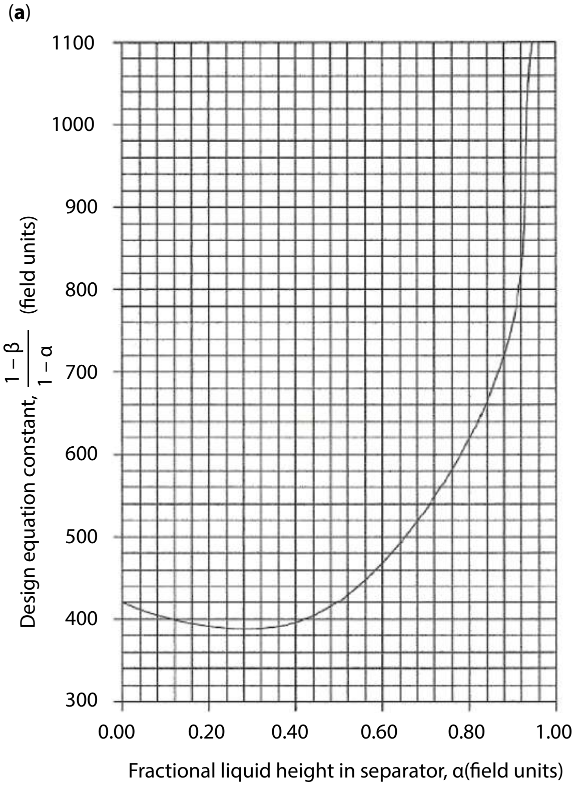


Figure 18.60a Gas capacity constraint design constant [(1 – β)/(1 – α)] vs. liquid height of a cylinder for a horizontal separator other than 50% full of liquid (field units) [79].

SI Units

$$L_{eff}d = 34.5\left(\frac{1-\beta}{1-\alpha}\right)\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l-\rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \tag{18.141}$$

where

$\left(\frac{1-\beta}{1-\alpha}\right)$ = design constant (Figure 18.60b)

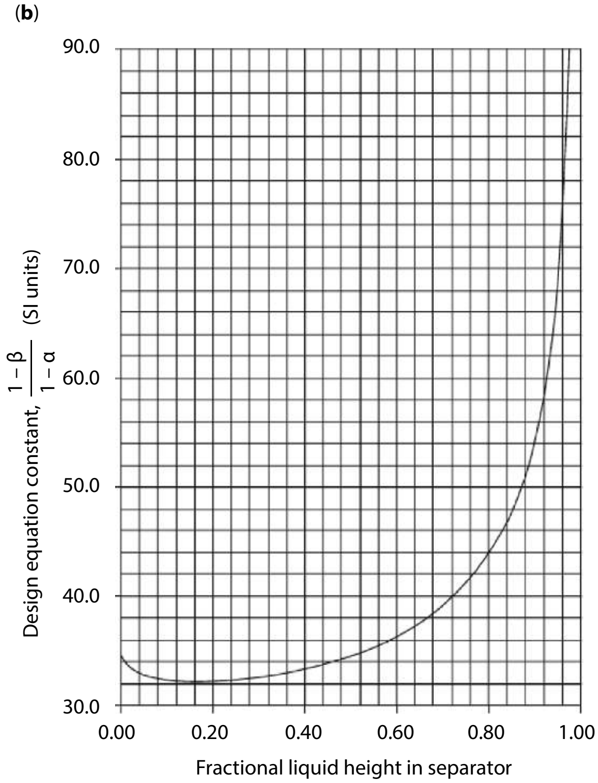


Figure 18.60b Gas capacity constraint design constant [(1 – β)/(1 – α)] vs. liquid height of a cylinder for a horizontal separator other than 50% full of liquid (SI units) [79].

Liquid Capacity Constraint

Field Units

$$d^2 Leff = \frac{t_r Q_l}{1.4\alpha} \tag{18.142}$$

where
α = design constant

If β is known, α can be determined from Figure 16.61a.

SI Units

$$d^2 Leff = \frac{21,221\, t_r Q_l}{\alpha} \tag{18.143}$$

where
α = design constant

If β is known, α can be determined from Figure 16.61b.

Sizing Vertical Separators

In vertical separators, a minimum diameter must be maintained to allow liquid droplets to separate from the vertically moving gas. The liquid retention time requirement specifies a combination of diameter and liquid volume height. Any diameter greater than the minimum required for gas capacity can be chosen. Figure 18.62 shows a

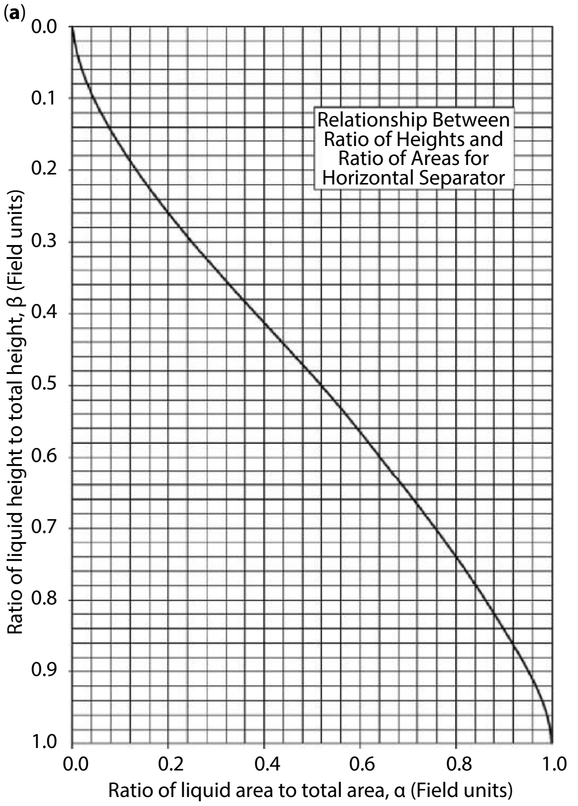


Figure 18.61a Liquid capacity constraint design constant—ratio of areas α vs. ratio of heights β for a horizontal separator other than 50% full of liquid (field units) [79].

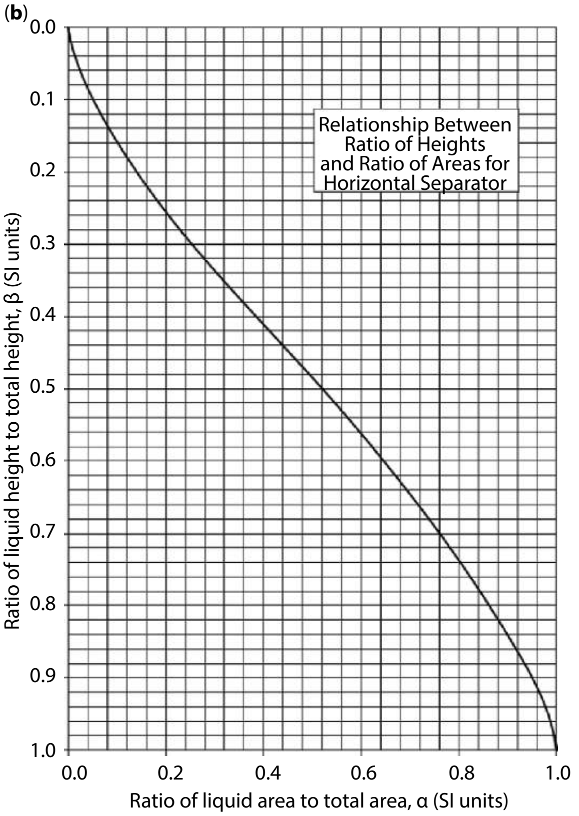


Figure 18.61b Liquid capacity constraint design constant—ratio of areas α vs. ratio of heights β for a horizontal separator other than 50% full of liquid (SI units) [79].

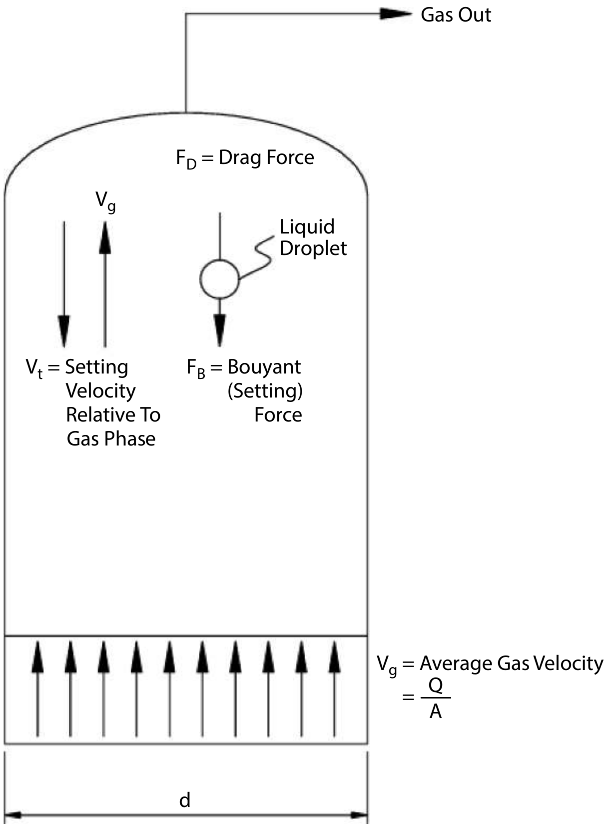


Figure 18.62 Model of a vertical separator [79].

model for a vertical separator and Arnold and Stewart provide the following guidelines for initial sizing of vertical two-phase separator [85].

Gas Capacity Constant

The principles of liquid droplets settling through a gas can be used to develop an equation to size a separator for a gas flow rate. When the retention time is equal to the time required for a droplet to settle to the liquid interface, the following equations are:

Field Units

$$d^2 = 5{,}040\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \tag{18.144}$$

SI Units

$$d^2 = 34{,}444\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \tag{18.145}$$

Eqs. 18.144 and 18.145 are derived as follows: for the droplets to fall, the gas velocity must be less than the terminal velocity of the droplet.

The terminal velocity, V_t is:

Field Units

$$V_t = 0.0119\left[\left(\frac{\rho_l - \rho_g}{\rho^g}\right)\frac{d_m}{C_D}\right]^{0.5} \tag{18.146}$$

SI Units

$$V_t = 0.0036\left[\left(\frac{\rho_l - \rho_g}{\rho^g}\right)\frac{d_m}{C_D}\right]^{0.5} \tag{18.147}$$

Determine the gas velocity, V_g:

where
Area, A = ft² (m²)
Diameter, D = ft. (m)
Diameter, d = in. (mm)
Volumetric flow rate, Q = ft³/s (m³/s)

Field Units

$$V_g = \frac{Q}{A_g} \tag{18.148}$$

$$\begin{aligned} A_g &= \frac{\pi D^2}{4} \\ &= \left(\frac{\pi}{4}\frac{d^2}{144}\right) = \frac{d^2}{183} \end{aligned} \tag{18.149}$$

Q_g is in MMscfd

$$\begin{aligned} Q &= Q_g \times 10^6\left(\frac{\text{scf}}{\text{MMscf}}\right)\left(\frac{\text{day}}{24\text{h}}\right)\left(\frac{\text{h}}{3600\text{s}}\right) \times \frac{14.7}{P} \times \frac{TZ}{520} \\ &= 0.327\frac{TZ}{P}Q_g \end{aligned} \tag{18.150}$$

$$V_g = \frac{\left(0.327\frac{TZ}{P}Q_g\right)(183)}{d^2} = 60\frac{TZQ_g}{Pd^2} \tag{18.151}$$

$$V_t = V_g \tag{18.152}$$

$$0.0119\left[\left(\frac{\rho_l-\rho_g}{\rho_g}\right)\frac{d_m}{C_D}\right]^{0.5}=\frac{60TZQ_g}{Pd^2} \tag{18.153}$$

$$d^2=5{,}040\frac{TZQ_g}{P}\left[\left(\frac{\rho_g}{\rho_l-\rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \tag{18.154}$$

SI Units

$$V_g=\frac{Q}{A_g} \tag{18.148a}$$

$$\begin{aligned}A_g&=\frac{\pi D^2}{4}\\&=\left(\frac{\pi}{4}\frac{d^2}{1000^2}\right)=7.855\times10^{-7}\,d^2\end{aligned} \tag{18.155}$$

$$Q_g = \text{scm/s}$$

$$Q=Q_g\times\frac{101.3}{P}\times\frac{TZ}{288.6}\left(\frac{h}{3600s}\right)=9.75\times10^{-5}\,\frac{TZ}{P}Q_g \tag{18.156}$$

$$V_g=\frac{\left(9.75\times10^{-5}\,\frac{TZ}{P}Q_g\right)}{7.855\times10^{-7}\,d^2}=124\frac{TZQ_g}{Pd^2} \tag{18.157}$$

$$V_t = V_g \tag{18.152}$$

$$\begin{aligned}0.0036\left[\left(\frac{\rho_l-\rho_g}{\rho_g}\right)\frac{d_m}{C_D}\right]^{0.5}&=124\frac{TZQ_g}{Pd^2}\\d^2=34{,}444\frac{TZQ_g}{P}&\left[\left(\frac{\rho_g}{\rho_l-\rho_g}\right)\frac{C_D}{d_m}\right]^{0.5}\end{aligned} \tag{18.158}$$

Liquid Capacity Constraint

Two-phase separators must be sized to provide some liquid retention time so the liquid can reach phase equilibrium with the gas. For a specified liquid flow rate and retention time, the following may be used to determine a vessel size.

Field Units

$$d^2h = \frac{t_r Q_l}{0.12} \tag{18.159}$$

SI Units

$$d^2h = \frac{t_r Q_l}{4.713 \times 10^{-7}} \tag{18.160}$$

where

h = height of the liquid volume, in. (mm)

Eqs. 18.159 and 18.160 are derived as follows:

Field Units

$$t = \frac{V}{Q} \tag{18.161}$$

$$V = \frac{\pi D^2 h}{(4)(12)} = \frac{\pi d^2 h}{4 \times 144 \times 12} = 4.55 \times 10^{-4} d^2 h \tag{18.162}$$

Q_l is in bpd

$$Q = Q_l \times 5.61 \left(\frac{ft^3}{barrel}\right)\left(\frac{day}{24h}\right)\left(\frac{h}{3600s}\right) = 6.49 \times 10^{-5} Q_l \tag{18.163}$$

$$t = \frac{V}{Q} = \frac{4.55 \times 10^{-4} d^2 h}{6.49 \times 10^{-5} Q_l} = 7.0 \frac{d^2h}{Q_l} \tag{18.164}$$

t_r is in min. = $60t_r$

$$d^2h = \frac{t_r Q_l}{0.12} \tag{18.165}$$

SI Units

$$t = \frac{V}{Q} \tag{18.161}$$

$$V = \frac{\pi D^2}{4} \times \frac{h}{1000} = \frac{\pi d^2 h}{4 \times 1000^2 \times 1000} = 7.854 \times 10^{-9} d^2 h \tag{18.166}$$

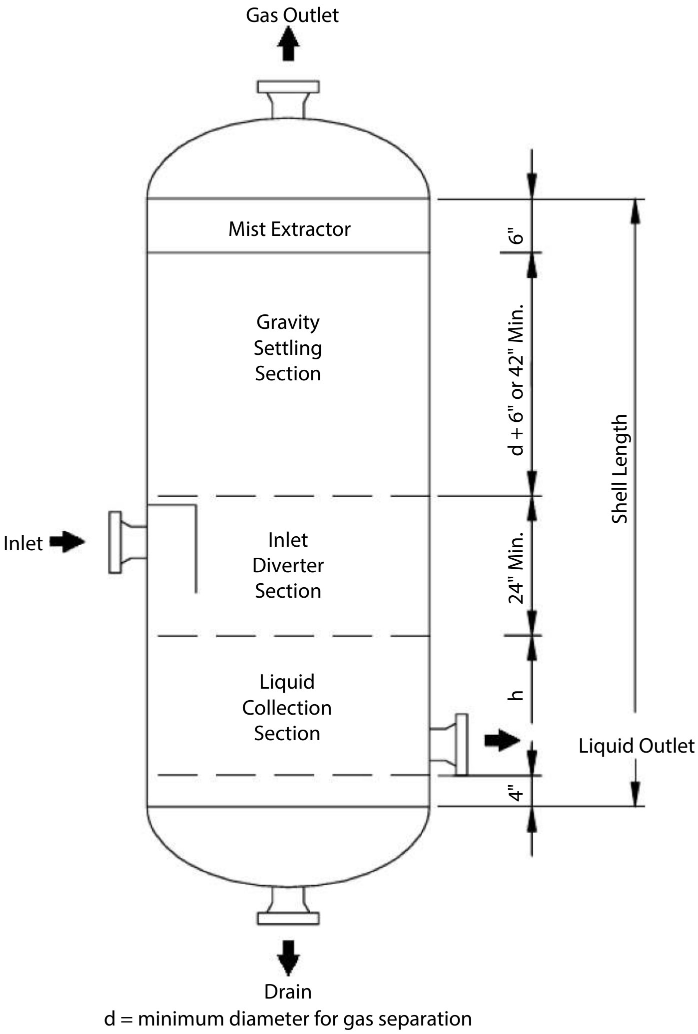
$$Q_l = m^3/h$$

$$Q = Q_l \times \left(\frac{h}{3600s}\right) = \frac{Q_l}{3600} \tag{18.167}$$

$$t = \frac{V}{Q} = \frac{7.853 \times 10^{-9}\, d^2\, h}{\dfrac{Q_l}{3600}} = 2.828 \times 10^{-5}\, \frac{d^2 h}{Q_l} \tag{18.168}$$

t_r is in min

$$d^2 h = \frac{t_r\, Q_l}{4.713 \times 10^{-7}} \tag{18.169}$$



d = minimum diameter for gas separation

Figure 18.63 Approximate seam-to-seam shell length for a vertical separator [79].

Seam-to Seam Length

The seam-to-seam length of vertical vessels may be estimated based on the diameter and liquid height. Allowance is allowed for the gas separation section and mist extractor and for any space below the water outlet as shown on Figure 18.63. The seam-to-seam length L_{ss} is determined by:

Field Units

$$L_{ss} = \frac{h + 76}{12} \text{ (for diameters} \le 36 \text{ in.)} \tag{18.170}$$

$$L_{ss} = \frac{h + d + 40}{12} \text{ (for diameters} > 36 \text{ in.)} \tag{18.171}$$

SI Units

$$L_{ss} = \frac{h + 1930}{1000} \text{ (for diameter} \le 194 \text{ mm)} \tag{18.172}$$

$$L_{ss} = \frac{h + d + 1016}{1000} \text{ (for diameter} > 194 \text{ mm)} \tag{18.173}$$

where

h = height of liquid level, in. (mm)
d = vessel ID, in. (mm)

The larger of the L_{ss} values from Eqs. 18.170–18.173 should be used.

Slenderness Ratio

In vertical separators where sizing is liquid dominated, it is generally common to choose slenderness ratios not greater than 4. This is to keep the height of the collection section to a reasonable level, and choices of between 3 and 4 are common, although height restriction may be that the slenderness ratio is lower. Generally, the larger the slenderness ratio, the less expensive is the vessel.

Procedure for Sizing Vertical Separators

The following steps are [79]:

1. Establish the design basis. This includes specifying the maximum and minimum flow rates, operating temperature and pressure, droplet size to be removed, and so on.
2. Use Eqs. 18.144 and 18.145 to determine the minimum required d. Any diameter larger than this value may be used.
3. For a selected d, Eqs. 18.159 and 18.160 may be used to determine h.
4. From d and h, the seam-to-seam length may be estimated using Eqs. 18.170–18.173. The larger value of L_{ss} should be used.
5. Check the slenderness ratio to determine if it is less than 4.
6. Make a final selection; always select a standard vessel size. Vessels with outside diameters up through 24 in. (600 mm) have nominal pipe dimensions. Vessels with outside diameters larger than 24 in. (600 mm) are rolled from plate with diameter increments of 6 in. (150 mm). The shell seam-to-seam length

is expanded in 2.5 ft (750 mm) segments and is usually from 5 to 10 ft (1500 to 3000 mm). Standard separator vessel sizes may be obtained from API 121J [75].

Example 18.10 Size a Vertical Two-Phase Separator (Field Units)

Given:

Gas flow rate:	10 MMscfd at 0.6 specific gravity
Oil flow rate:	2000 bpd at 40°API
Operating pressure	1000 psia
Operating temperature	60°F
Droplet size removal	140 μm
Retention time	3 min
Compressibility factor, Z	0.84
Viscosity, μ	0.013 cP
Assume C_D	0.34

Solution

1. Calculate C_D

$$\rho_l = 62.4\left[\left(\frac{141.5}{(131.5+40)}\right)\right] = 51.5\frac{\text{lb}}{\text{ft}^3}$$

$$\rho_g = 2.70\frac{(\text{SpGr})(\text{P})}{\text{TZ}} = 2.70\frac{(0.6)(1{,}000)}{(520)(0.84)} = 3.71\frac{\text{lb}}{\text{ft}^3}$$

$$d_m = 140\ \mu\text{m}$$

The terminal velocity, V_t from Eq. 18.146 is:

$$V_t = 0.0119\left[\left(\frac{51.5-3.71}{3.71}\right)\frac{140}{0.34}\right]^{0.5} = 0.867\ \text{ft/s}$$

Reynolds number, Re:

$$\text{Re} = 0.0049\frac{\rho_g\, d_p\, V_t}{\mu}$$

$$\text{Re} = 0.0049\left[\frac{(3.71)(140)(0.867)}{0.013}\right] = 169.54$$

Drag coefficient, C_D:

$$C_D = \frac{24}{Re} + \frac{3}{Re^{0.5}} + 0.34$$
$$C_D = \frac{24}{169.54} + \frac{3}{(169.54)^{0.5}} + 0.34 = 0.712$$

Repeat using $C_D = 0.712$

$$V_t = 0.599 \text{ ft/s}$$
$$Re = 117$$
$$C_D = 0.822$$

Repeat:

$$V_t = 0.556 \text{ ft/s}$$
$$Re = 110$$
$$C_D = 0.844$$

Repeat:

$$V_t = 0.548 \text{ ft/s}$$
$$Re = 108$$
$$C_D = 0.851$$

Repeat:

$$V_t = 0.545 \text{ ft/s}$$
$$Re = 108$$
$$C_D = 0.854 \text{ o.k.}$$

2. Gas Capacity Constraint

From Eq. 18.144

$$d^2 = 5040\left[\frac{(520)(0.84)(10)}{1{,}000}\right]\left[\left(\frac{3.71}{51.5-3.71}\right)\frac{0.854}{140}\right]^{0.5}.$$
$$d = 21.9 \text{ in.}$$

3. Liquid Capacity Constraint

From Eq. 18.119

Table 18.17 Vertical separator Example 18.10: Diameter vs. length for liquid capacity constraint.

t_r (min)	d (in.)	h (in.)	L_{ss} (ft)	$S_R\left(\frac{12L_{ss}}{d}\right)$
3	24	86.8	13.6	6.8
	30	55.6	11.0	4.4
	36	38.6	9.6	3.2
	42	28.3	8.7	2.5
	48	21.7	8.1	2.0
2	24	57.9	11.2	5.6
	30	37.0	9.4	3.8
	36	25.7	8.5	2.8
	42	18.9	7.9	2.3
1	24	28.9	8.7	4.4
	30	18.5	7.9	3.2
	36	12.9	7.4	2.5

4. Calculate combinations of d and h for various t_r (see Table 18.17).
5. Calculate seam-to-seam length from Eq. 18.170 or Eq. 18.171 (see Table 18.17).
6. Calculate slenderness ratio: 12 L_{ss}/d. Choices in the range of 3–4 are common (see Table 18.17).
7. Choose a reasonable size with a diameter greater than that determined by the gas capacity. A 36 in. diameter by 10 ft seam-to-seam separator provides slightly more than 3 min. retention time with a diameter greater than 21.8 in. and a slenderness ratio of 3.2.

The Excel spreadsheet Example 18.10.xlsx shows the calculations of Example 18.10 and the table below shows the results of the spreadsheet calculations.

Sizing Calculation of a vertical two-phase separator (Field Units)		
Gas flow rate	10	MMscfd
Oil flow rate	2000	bpd
SG of oil	40	°API
SG of gas	0.6	
Operating temperature	60	°F
Operating pressure	1000	psia
Droplet size removal	140	micron
Retention time, min	3	min.
Compressibility factor, Z	0.84	
Viscosity, μ	0.013	cP

Assume Drag coefficient, C_D	0.34	
Reynolds number	108	
Calculated Drag coefficient, C_D	0.8509	
Terminal velocity, V_t	0.548	ft/s
Vessel diameter, d	21.9	in.

Example 18.11 Size a Vertical Two-Phase Separator (SI Units)

Given:

Gas flow rate	11,803 scm/h at 0.6 specific gravity
Oil flow rate	317.6 m^3/h at 40°API
Operating pressure	69,000 kPa
Operating temperature	15.60°C
Droplet size removal	140 μm
Retention time	3 min
Compressibility factor, Z	0.84
Viscosity, μ	0.013 cP
Assume C_D	0.34

Solution

1. Calculate C_D

$$\rho_l = 1000\left[\left(\frac{141.5}{(131.5+40)}\right)\right] = 825\,\frac{\text{kg}}{\text{m}^3}$$

$$\rho_g = 3.492\frac{(\text{SpGr})(\text{P})}{\text{TZ}} = 3.492\frac{(0.6)(6900)}{(288.6)(0.84)} = 59.6\,\frac{\text{kg}}{\text{m}^3}$$

$$d_m = 140\ \mu\text{m}$$

The terminal velocity, V_t from Eq. 18.147 is:

$$V_t = 0.0036\left[\left(\frac{825-59.6}{59.6}\right)\frac{140}{0.34}\right]^{0.5} = 0.2618\ \text{m/s}$$

Reynolds number, Re:

$$\text{Re} = 0.001\frac{\rho_g\, d_p\, V_t}{\mu}$$

$$\text{Re} = 0.001\left[\frac{(59.6)(140)(0.2618)}{0.013}\right] = 168$$

Drag coefficient, C_D:

$$C_D = \frac{24}{Re} + \frac{3}{Re^{0.5}} + 0.34$$

$$C_D = \frac{24}{168} + \frac{3}{(168)^{0.5}} + 0.34 = 0.714$$

Repeat using $C_D = 0.714$

$V_t = 0.1812$ m/s

Re = 116

$C_D = 0.86$

Repeat:

$V_t = 0.1686$ m/s

Re = 108

$C_D = 0.851$—o.k.

2. Gas Capacity Constraint
From Eq. 18.145

$$d^2 = 34,444\left[\frac{(288.8)(0.84)(11803)}{6900}\right]\left[\left(\frac{59.6}{825-59.6}\right)\frac{0.851}{140}\right]^{0.5} = 557.5\,\text{mm.}$$

Table 18.18 Vertical separator Example 18.10: Diameter vs. length for liquid capacity constraint.

t_r (min)	d (mm.)	h (mm.)	L_{ss} (m)	$S_R = \left(\frac{1000L_{ss}}{d}\right)$
3	609.6	5440	7.1	11.6
	762	3482	5.3	7
	914.4	2418	4.3	4.7
	1066.8	1776	3.9	3.7
	1219.2	1360	3.6	3
2	609.6	2627	5.3	8.7
	762	2321	4.1	5.4
	914.4	1612	3.5	3.8
	1066.8	1184	3.3	3.1
1	609.6	1813	3.4	5.6
	762	1161	2.9	3.8
	914.4	806	2.7	3

3. Liquid Capacity Constraint
   From Eq. 18.160

$$d^2h = \frac{t_r\,Q_l}{4.713 \times 10^{-7}}$$

4. Calculate combinations of d and h for various t_r (see Table 18.18).
5. Calculate seam-to-seam length from Eq. 18.172 or Eq. 18.173 (see Table 18.18)

$$L_{ss} = \frac{h + 1930}{1000} \text{ for (diameters} \le 194 \text{ mm)}$$

$$L_{ss} = \frac{h + d + 1016}{1000} \text{ for (diameter} > 194 \text{ mm)}$$

6. Calculate slenderness ratio: $1000L_{ss}/d$. Choices in the range of 3–4 are common (see Table 18.18).
7. Choose a reasonable size with a diameter greater than that determined by the gas capacity. A 914 mm diameter by 3 m seam-to-seam separator provides slightly more than 3 min. retention time with a diameter greater than 557.5 mm and a slenderness ratio of 4.7.

The Excel spreadsheet Example 18.11.xlsx shows the calculations of Example 18.11 and the table below shows the results of the spreadsheet calculations.

Sizing Calculation of a vertical two-phase separator (S.I. Units)		
Gas flow rate	11803	scm/h
Oil flow rate	317.6	m^3/h
SG of oil	40	°API
SG of gas	0.6	
Operating temperature	15.6	°C
Operating pressure	6900	kPa
Droplet size removal	140	micron
Retention time, min	3	min.
Compressibility factor, Z	0.84	
Viscosity, μ	0.013	cP
Assume Drag coefficient, C_D	0.34	
Reynolds number	106	
Calculated Drag coefficient, C_D	0.858	
Terminal velocity, V_t	0.165	m/s
Vessel diameter, d	558.7	mm

Example 18.12 Size a Horizontal Two-Phase Separator (Field Units)

Given:

Gas flow rate	10 MMscfd at 0.6 specific gravity
Oil flow rate	2000 bpd at 40°API
Operating pressure	1000 psia
Operating temperature	60°F
Droplet size removal	140 μm
Retention time	3 min
Compressibility factor, Z	0.84
Viscosity, μ	0.013 cP
Assume C_D	0.34

Solution

1. Calculate C_D (same as Examples 18.10 and 18.11)

$$C_D = 0.851$$

2. Gas capacity constraint from Eq. 18.105

$$dL_{eff} = 420\left[\frac{(520)(0.84)(10)}{1{,}000}\right]\left[\left(\frac{3.71}{51.5-3.71}\right)\frac{0.851}{140}\right]^{0.5}$$
$$= 39.85$$

3. Liquid capacity constraint from Eq. 18.124

$$d^2 L_{eff} = \frac{t_r Q_l}{0.7}$$

4. Calculate combinations of d and L_{ss} for gas and liquid capacity.
5. Calculate seam-to-seam length from Eq. 18.137 for various d (Table 18.19).

$$L_{ss} = \frac{4}{3} L_{eff}$$

6. Calculate slenderness ratio S_R from Eq. 18.138. Choices in the range of 3–4.

$$S_R = \frac{12 L_{ss}}{d}$$

7. Choose a reasonable size with a diameter and length combination above both the gas capacity and the liquid capacity constraint lines. A 36-in. × 10 ft separator provides about 3 min. retention time.

Table 18.19 Horizontal separator diameter vs. length.

d (in.)	Gas L_{eff} (ft)	Liquid L_{eff} (ft)	L_{ss} (ft)	$S_R = \frac{12L_{ss}}{d}$
16	2.5	33.5	44.7	33.5
20	2	21.4	28.5	17.1
24	1.7	14.9	19.9	10
30	1.3	9.5	12.7	5.1
36	1.1	6.6	8.8	2.9
42	1	4.9	6.5	1.9
48	0.8	3.7	4.9	1.2
54	0.7	2.9	3.9	0.9
60	0.7	2.4	3.2	0.6

The Excel spreadsheet Example 18.12.xlsx shows the calculations of Example 18.12, and the table below shows the results of the spreadsheet calculations.

Sizing Calculation of a horizontal two-phase separator (Field Units)		
Gas flow rate	10	MMscfd
Oil flow rate	2000	bpd
SG of oil	40	°API
SG of gas	0.6	
Operating temperature	60	°F
Operating pressure	1000	psia
Droplet size removal	140	micron
Retention time, min	3	min.
Compressibility factor, Z	0.84	
Viscosity, μ	0.013	cP
Assume Drag coefficient, C_D	0.34	
Reynolds number	108	
Calculated Drag coefficient, C_D	0.8509	
Terminal velocity, V_t	0.548	ft/s
Vessel diameter, d	36	in.

Example 18.13 Sizing a Horizontal Two-Phase Separator (SI Units)

Given:

Gas flow rate	11,803 scf/h at 0.6 specific gravity
Oil flow rate	13.25 m^3/h at 40°API

Operating pressure	6900 kPa
Operating temperature	15.6°C
Droplet size removal	140 μm
Retention time	3 min
Compressibility factor, Z	0.84
Viscosity, μ	0.013 cP
Assume C_D	0.34

Solution

1. Calculate C_D (same as Examples 18.10 and 18.11)

$$C_D = 0.851$$

Densities of oil and gas are:

$$\rho_l = 1000\left[\frac{141.5}{(131.5+40)}\right] = 825\frac{kg}{m^3}$$

$$\rho_g = 3.492\frac{(SpGr)(P)}{TZ} = 3.492\frac{(0.6)(6900)}{(288.75)(0.84)} = 59.6\frac{kg}{m^3}$$

2. Gas capacity constraint from Eq. 18.106

$$dL_{eff} = 34.5\left[\frac{(288.75)(0.84)(11803)}{6900}\right]\left[\left(\frac{59.6}{825-59.6}\right)\frac{0.8578}{140}\right]^{0.5}$$
$$= 312.66$$

3. Liquid capacity constraint from Eq. 18.125

$$d^2L_{eff} = 42445\ t_r\ Q_l$$

4. Calculate combinations of d and L_{ss} for gas and liquid capacity.
5. Calculate seam-to-seam length from Eq. 18.137 for various d (Table 18.20).

$$L_{ss} = \frac{4}{3}L_{eff}$$

6. Calculate slenderness ratio S_R from Eq. 18.138. Choices in the range of 3–4.

$$S_R = \frac{1000\,L_{ss}}{d}$$

Table 18.20 Horizontal separator, diameter vs. length.

d (mm)	Gas L_{eff} (m)	Liquid L_{eff} (m)	L_{ss} (m)	$S_R = \frac{1000L_{ss}}{d}$
406.4	0.77	10.22	13.62	33.5
508	0.62	6.54	8.7	17.1
609.6	0.51	4.54	6.1	10
762	0.41	2.91	3.9	5.1
914.4	0.39	2.02	2.7	3.0
1066.8	0.29	1.48	2.0	1.9
1219.2	0.26	1.14	1.5	1.2

7. Choose a reasonable size with a diameter and length combination above both the gas capacity and the liquid capacity constraint lines. A 914 mm by 3 m separator provides about 3 min retention time.

The Excel spreadsheet Example 18.13.xlsx shows the calculations of Example 18.13, and the table below shows the results of the spreadsheet calculations.

Sizing Calculation of a horizontal two-phase separator (S.I. Units)

Gas flow rate	11803	scm/h
Oil flow rate	317.6	m^3/h
SG of oil	40	°API
SG of gas	0.6	
Operating temperature	15.6	°C
Operating pressure	6900	kPa
Droplet size removal	140	micron
Retention time, min	3	min.
Compressibility factor, Z	0.84	
Viscosity, μ	0.013	cP
Assume Drag coefficient, C_D	0.34	
Reynolds number	106	
Calculated Drag coefficient, C_D	0.8509	
Terminal velocity, V_t	0.168	m/s
Vessel diameter, d	914.4	mm

A Case Study

A wet gas consists of methane (CH_4), ethane (C_2H_6), nitrogen (N_2), carbon dioxide (CO_2), and water (H_2O) at a temperature of 45°C, pressure at 350 kPa and flow rate at 399.2 kg mole/h is to be separated in a horizontal flat dish separator of a diameter of 4.572 m and length of 16 m having a boot. Use UniSim Design simulation software to size the horizontal separator. The compositions of the wet gas are as follows:

Composition	Molar flow (kg mole/h)
Methane (CH_4)	382.75
Ethane (C_2H_6)	0.07984
Propane (C_3H_8)	0.0000
i-Butane (iC_4H_{10})	0.0000
n-Butane (C_4H_{10})	0.0000
Nitrogen (N_2)	7.1856
Carbon dioxide (CO_2)	0.63872
Hydrogen sulfide (H_2S)	0.0000
Water (H_2O)	8.5429
Total	399.20

Solution

UniSim Design R460.1 simulation software was employed to size the horizontal separator and the Property Package of Soave–Redlich–Kwong (SRK). The following screen shots illustrate the input data and final simulation results can be accessed from the simulation file (Wet Gas gathering Separator.usc). Figures 18.64–18.71 show the screen shots of the input data and the process flow diagram of the horizontal Wet Gas separator and Table 18.21 shows the results of the simulation.

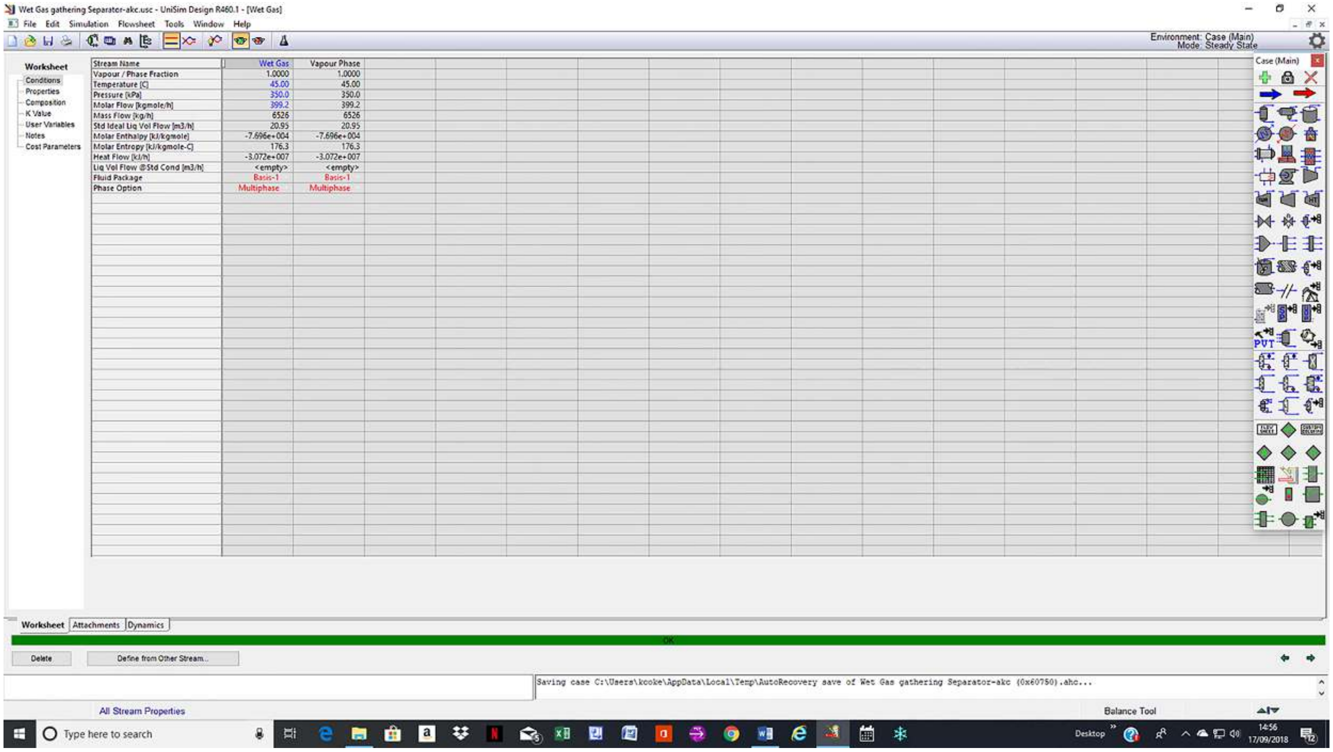
The results of the separator from UniSim simulation design software are shown in Table 18.21.

Three-Phase Separators

Three-phase separators handle two immiscible liquids (oil and water) and a gas. The design of the liquid–gas separation is the same as that for the two-phase separators, except that in this case the densities of both liquids (oil and water) have to be considered. Usually, a certain separation of oil from the water layer and water from the oil layer is to be supplied to act as a guide to the design of the liquid–liquid separation portion. A certain size of the droplets of the distributed phase (500 μm) in the continuous phase is assumed.

Figure 18.72 illustrates a horizontal three-phase separator with interface level control and weir. The fluid enters the separator and hits an inlet inverter, as the sudden change in momentum does the initial gross separation of liquid and vapor. The inlet diverter contains a down-comer that directs the liquid flow below the oil-water interface. This forces the inlet mixture of oil and water to mix with the water continuous phase in the bottom of the vessel and rise through the oil–water interface. This process is referred to as "water washing" and it promotes the coalescence of water droplets, which are entrained in the oil continuous phase.

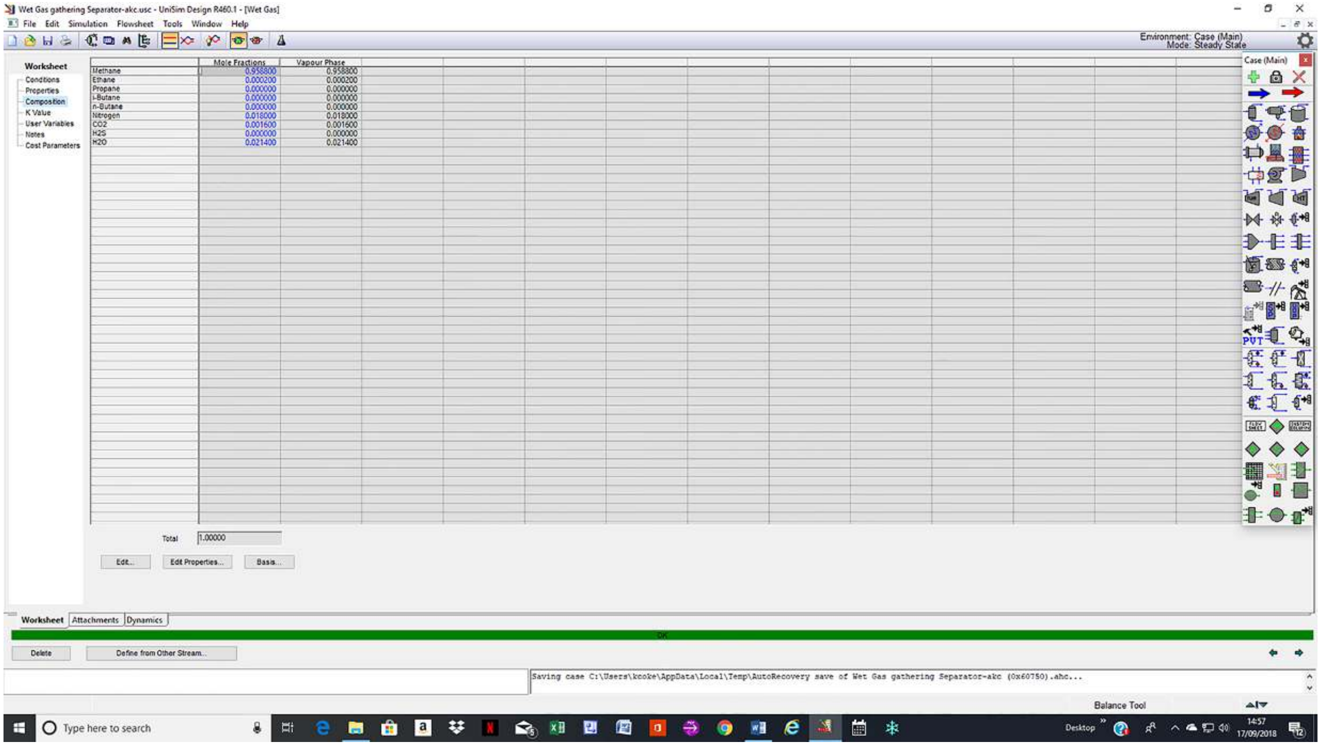
Calculations are carried out to determine whether the droplet will reach the interface layer in the time that it will take for the travel using settling/buoyancy phenomena. Furthermore, calculations are cross-checked with minimum residence times required for oil and water to achieve the required separation. American Petroleum Institute (API 12J) standard recommends the following residence times based on the relative density of oil. Residence times are usually fixed using more rigorous calculations accounting for settling/buoyancy effects [75]. Figure 18.73 shows a



Stream Name	Wet Gas	Vapour Phase
Vapour / Phase Fraction	1.0000	1.0000
Temperature [C]	45.00	45.00
Pressure [kPa]	350.0	350.0
Molar Flow [kgmole/h]	399.2	399.2
Mass Flow [kg/h]	6526	6526
Std Ideal Liq Vol Flow [m3/h]	20.95	20.95
Molar Enthalpy [kJ/kgmole]	-7.696e+004	-7.696e+004
Molar Entropy [kJ/kgmole-C]	176.3	176.3
Heat Flow [kJ/h]	-3.072e+007	-3.072e+007
Liq Vol Flow @Std Cond [m3/h]	<empty>	<empty>
Fluid Package	Basis-1	Basis-1
Phase Option	Multiphase	Multiphase

```
Saving case C:\Users\kcoke\AppData\Local\Temp\AutoRecovery save of Wet Gas gathering Separator-akc (0x60750).ahc...
```

Figure 18.64 A snapshot of the Worksheet window showing the Conditions page (courtesy of Honeywell Process Solution. UniSim Design R460.1, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).

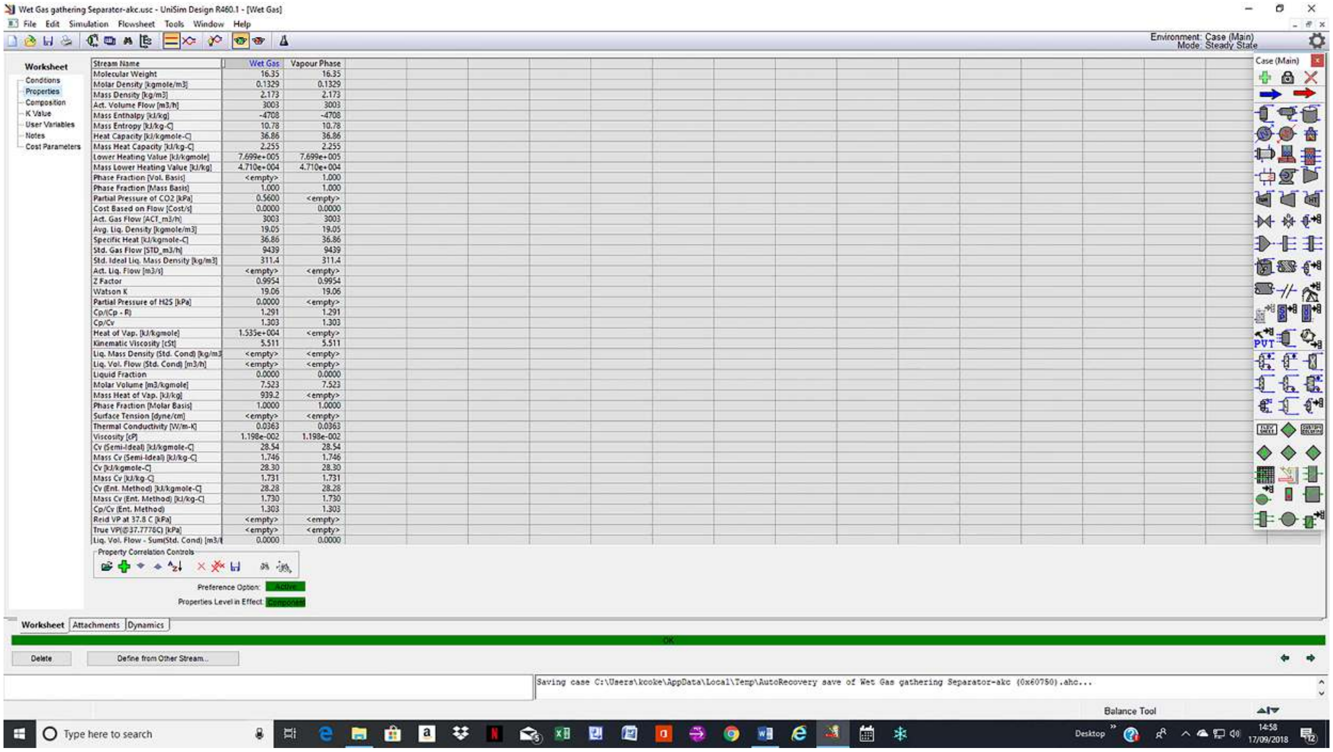


	Mole Fractions	Vapour Phase
Methane	0.958800	0.958800
Ethane	0.000200	0.000200
Propane	0.000000	0.000000
i-Butane	0.000000	0.000000
n-Butane	0.000000	0.000000
Nitrogen	0.018000	0.018000
CO2	0.001600	0.001600
H2S	0.000000	0.000000
H2O	0.021400	0.021400

Total 1.00000

```
Saving case C:\Users\kcoke\AppData\Local\Temp\AutoRecovery save of Wet Gas gathering Separator-akc (0x60750).ahc...
```

Figure 18.65 A snapshot of the Worksheet window showing the Compositions page (courtesy of Honeywell Process Solution. UniSim Design R460.1, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).


Wet Gas gathering Separator-akc.usc - UniSim Design R460.1 - [Wet Gas]

Stream Name	Wet Gas	Vapour Phase
Molecular Weight	16.35	16.35
Molar Density [kgmole/m3]	0.1329	0.1329
Mass Density [kg/m3]	2.173	2.173
Act. Volume Flow [m3/h]	3003	3003
Mass Enthalpy [kJ/kg]	-4708	-4708
Mass Entropy [kJ/kg-C]	10.78	10.78
Heat Capacity [kJ/kgmole-C]	36.86	36.86
Mass Heat Capacity [kJ/kg-C]	2.255	2.255
Lower Heating Value [kJ/kgmole]	7.699e+005	7.699e+005
Mass Lower Heating Value [kJ/kg]	4.710e+004	4.710e+004
Phase Fraction [Vol. Basis]	<empty>	1.000
Phase Fraction [Mass Basis]	1.000	1.000
Partial Pressure of CO2 [kPa]	0.5600	<empty>
Cost Based on Flow [Cost/s]	0.0000	0.0000
Act. Gas Flow [ACT_m3/h]	3003	3003
Avg. Liq. Density [kgmole/m3]	19.05	19.05
Specific Heat [kJ/kgmole-C]	36.86	36.86
Std. Gas Flow [STD_m3/h]	9439	9439
Std. Ideal Liq. Mass Density [kg/m3]	311.4	311.4
Act. Liq. Flow [m3/s]	<empty>	<empty>
Z Factor	0.9954	0.9954
Watson K	19.06	19.06
Partial Pressure of H2S [kPa]	0.0000	<empty>
Cp/(Cp - R)	1.291	1.291
Cp/Cv	1.303	1.303
Heat of Vap. [kJ/kgmole]	1.535e+004	<empty>
Kinematic Viscosity [cSt]	5.511	5.511
Liq. Mass Density (Std. Cond) [kg/m3	<empty>	<empty>
Liq. Vol. Flow (Std. Cond) [m3/h]	<empty>	<empty>
Liquid Fraction	0.0000	0.0000
Molar Volume [m3/kgmole]	7.523	7.523
Mass Heat of Vap. [kJ/kg]	939.2	<empty>
Phase Fraction [Molar Basis]	1.0000	1.0000
Surface Tension [dyne/cm]	<empty>	<empty>
Thermal Conductivity [W/m-K]	0.0363	0.0363
Viscosity [cP]	1.198e-002	1.198e-002
Cv (Semi-Ideal) [kJ/kgmole-C]	28.54	28.54
Mass Cv (Semi-Ideal) [kJ/kg-C]	1.746	1.746
Cv [kJ/kgmole-C]	28.30	28.30
Mass Cv [kJ/kg-C]	1.731	1.731
Cv (Ent. Method) [kJ/kgmole-C]	28.28	28.28
Mass Cv (Ent. Method) [kJ/kg-C]	1.730	1.730
Cp/Cv (Ent. Method)	1.303	1.303
Reid VP at 37.8 C [kPa]	<empty>	<empty>
True VP(@37.7776C) [kPa]	<empty>	<empty>
Liq. Vol. Flow - Sum(Std. Cond) [m3/	0.0000	0.0000

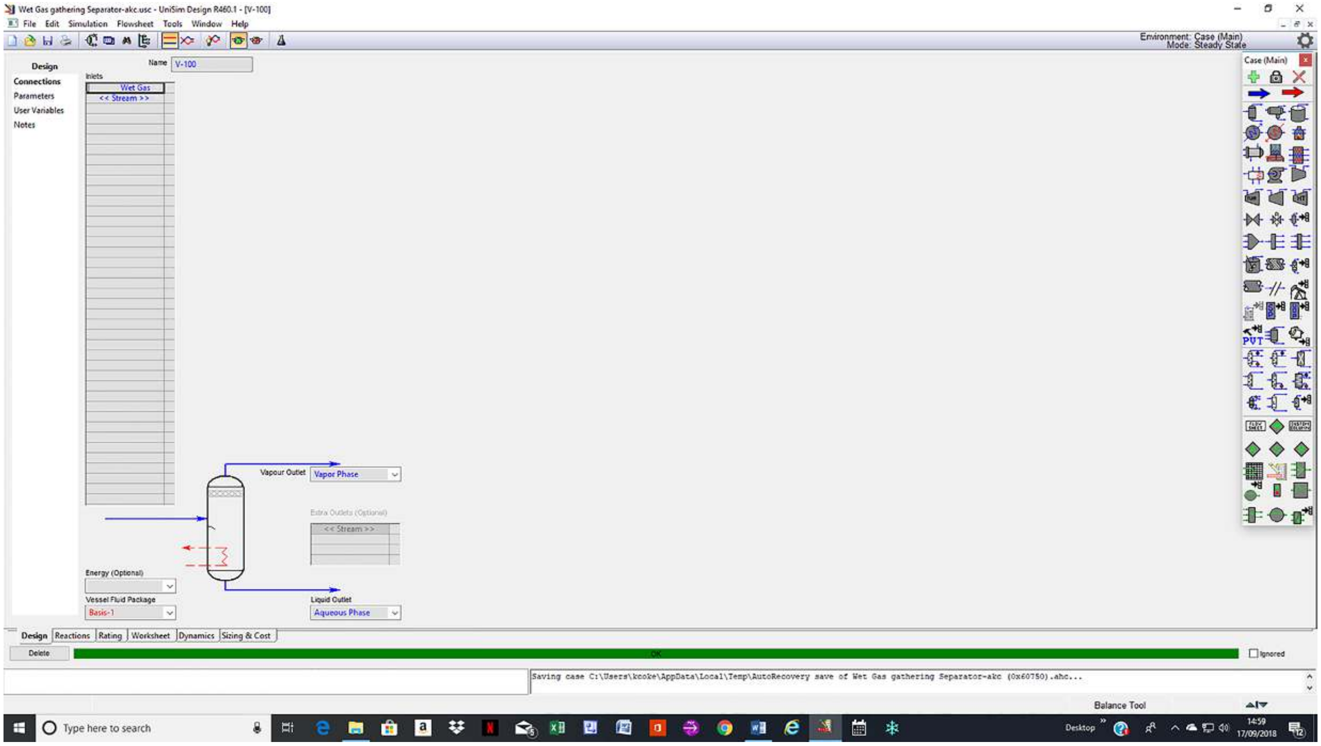
Preference Option: Active

Properties Level in Effect: Component

Environment: Case (Main)
Mode: Steady State

```
Saving case C:\Users\kcoke\AppData\Local\Temp\AutoRecovery save of Wet Gas gathering Separator-akc (0x60750).ahc...
```

Figure 18.66 A snapshot of the Worksheet window showing the Properties page (courtesy of Honeywell Process Solution. UniSim Design R460.1, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).


Wet Gas gathering Separator-akc.usc - UniSim Design R460.1 - [V-100]

Name: V-100

Inlets: Wet Gas

Vapour Outlet: Vapor Phase

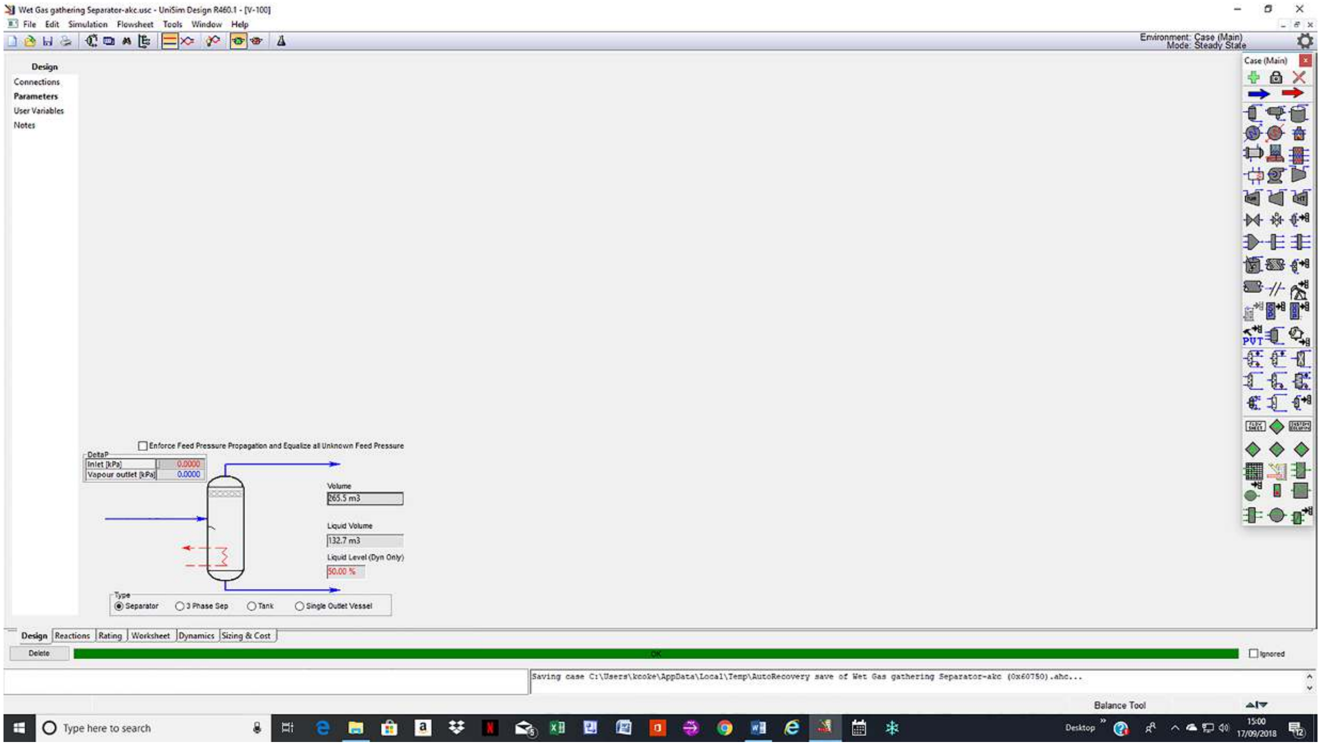
Liquid Outlet: Aqueous Phase

Vessel Fluid Package: Basis-1

Environment: Case (Main)
Mode: Steady State

```
Saving case C:\Users\kcoke\AppData\Local\Temp\AutoRecovery save of Wet Gas gathering Separator-akc (0x60750).ahc...
```

Figure 18.67 A snapshot of the Design window showing the Connections page (courtesy of Honeywell Process Solution. UniSim Design R460.1, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



☐ Enforce Feed Pressure Propagation and Equalize all Unknown Feed Pressure

DeltaP

Inlet [kPa]	0.0000
Vapour outlet [kPa]	0.0000

Volume: 265.5 m3

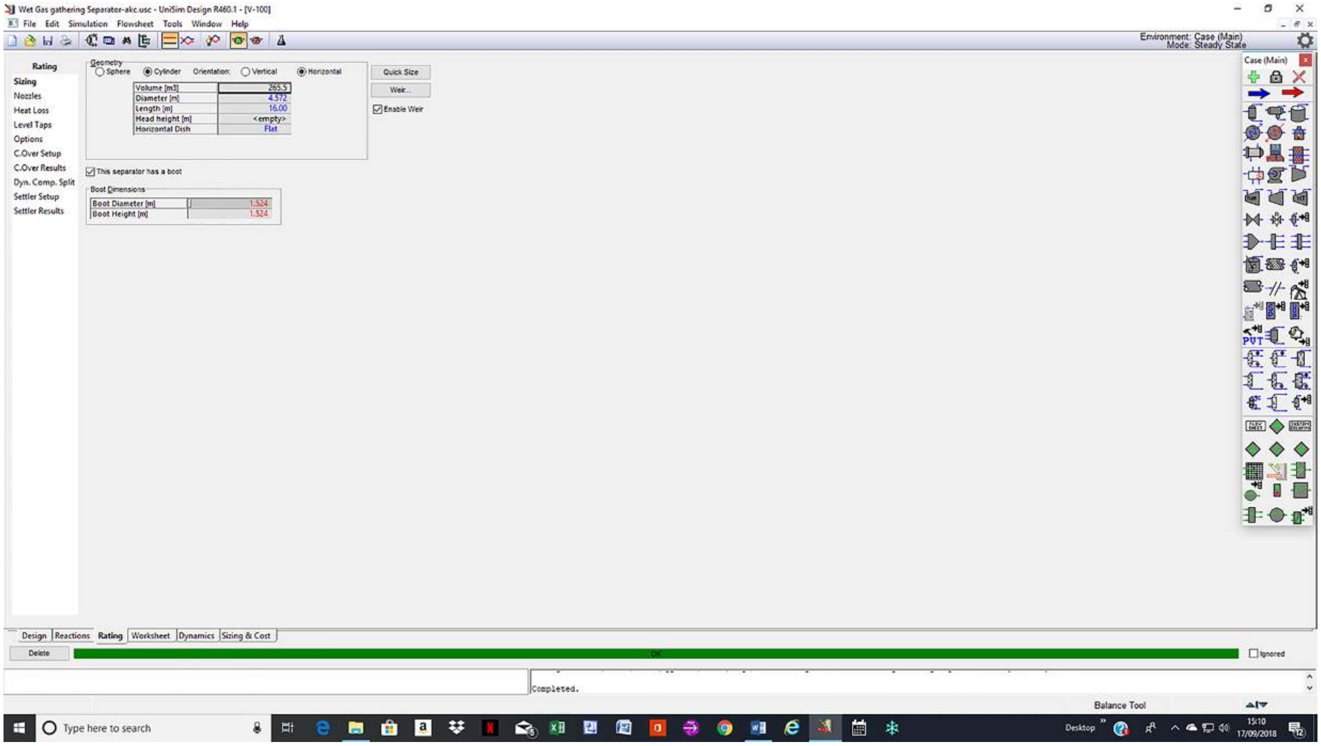
Liquid Volume: 132.7 m3

Liquid Level (Dyn Only): 50.00 %

Type: ● Separator ○ 3 Phase Sep ○ Tank ○ Single Outlet Vessel

```
Saving case C:\Users\kcoke\AppData\Local\Temp\AutoRecovery save of Wet Gas gathering Separator-akc (0x60750).ahc...
```

Figure 18.68 A snapshot of the Design window showing the Parameters page (courtesy of Honeywell Process Solution. UniSim Design R460.1, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



Geometry: ○ Sphere ● Cylinder Orientation: ○ Vertical ● Horizontal

Volume [m3]	265.5
Diameter [m]	4.572
Length [m]	16.00
Head height [m]	<empty>
Horizontal Dish	Flat

☑ Enable Weir

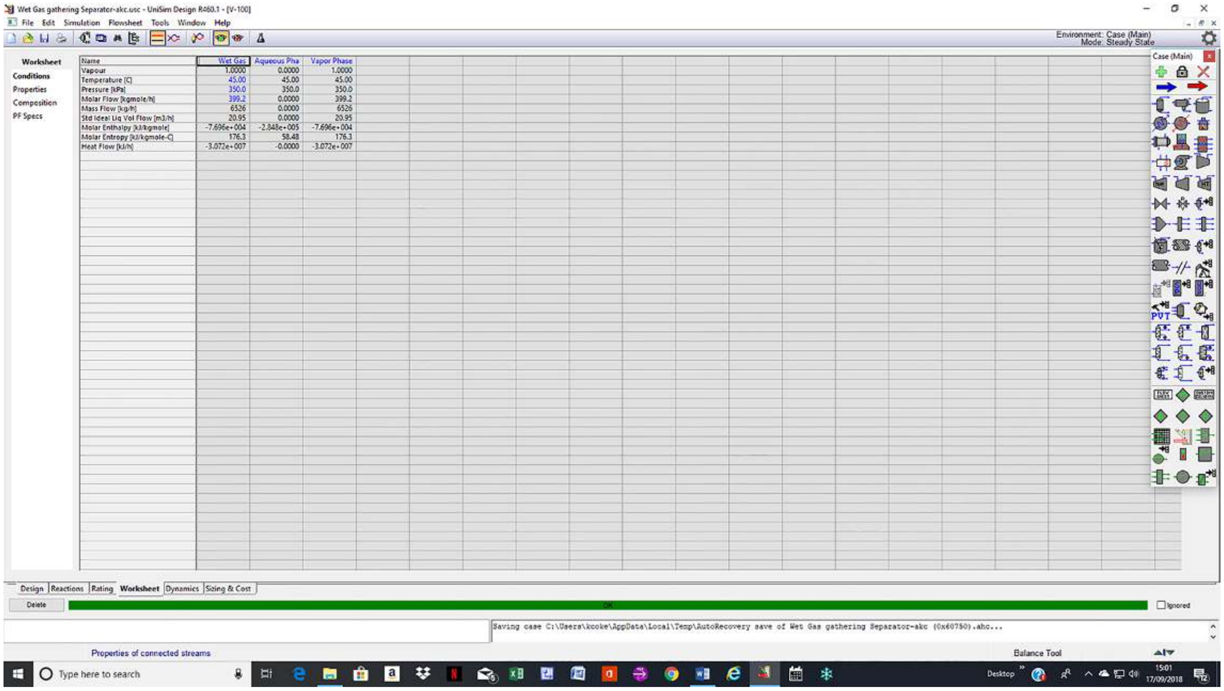
☑ This separator has a boot

Boot Dimensions

Boot Diameter [m]	1.524
Boot Height [m]	1.524

```
Completed.
```

Figure 18.69 A snapshot of the Rating window showing the Sizing page (courtesy of Honeywell Process Solution. UniSim Design R460.1, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



Name	Wet Gas	Aqueous Pha	Vapor Phase
Vapour	1.0000	0.0000	1.0000
Temperature [C]	45.00	45.00	45.00
Pressure [kPa]	350.0	350.0	350.0
Molar Flow [kgmole/h]	399.2	0.0000	399.2
Mass Flow [kg/h]	6526	0.0000	6526
Std Ideal Liq Vol Flow [m3/h]	20.95	0.0000	20.95
Molar Enthalpy [kJ/kgmole]	-7.696e+004	-2.848e+005	-7.696e+004
Molar Entropy [kJ/kgmole-C]	176.3	58.48	176.3
Heat Flow [kJ/h]	-3.072e+007	-0.0000	-3.072e+007

```
Saving case C:\Users\kcoke\AppData\Local\Temp\AutoRecovery save of Wet Gas gathering Separator-akc (0x60750).ahc...
```

Figure 18.70 A snapshot of the Worksheet window showing the Conditions page (courtesy of Honeywell Process Solution. UniSim Design R460.1, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).

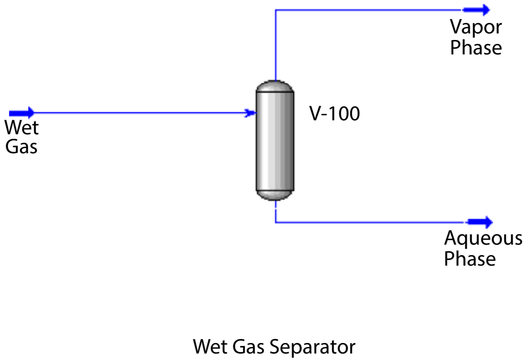


Figure 18.71 A snapshot of Process flow diagram of Wet Gas Separator (courtesy of Honeywell Process Solution. UniSim Design R460.1, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).

typical three-phase horizontal separator and Figure 18.74 illustrates the configurations and notations of the horizontal three-phase separator.

Oil relative density	Residence time, min.
Below 0.85	3–5
Above 0.85–100°F and above	5–10
80–100°F	10–20
60–80°F	20–30

Table 18.21 Results of process engineering design sizing using UniSim Design R460.1.

	Sizing value	**Simulation value**
Orientation	Vertical	
Type	VVLS	
Head Type	Elliptical	
Diameter, m	1.2	4.572
Tangent length, m	2.0	16.0
Liquid particle diameter, m	0.250	
Heavy liquid particle diameter, m	0.125	
Vapor particle diameter, m	0.1750	
Vapor flow, m^3/h	3003	
Heavy liquid flow, m^3/h	0	
Total liquid flow, m^3/h	0.2271	
Vapor density, kg/m^3	2.173	
Light liquid density, kg/m^3	800.9	
Heavy liquid density, kg/m^3	999.6	
Surge time (HLL < - > LLL), s	120.0	
Residence time, (NLL < - > Empty), s	6.0	
Vert Pctliq Full, %	50.0	
HorizPctLiq Full, %	66.0	
Vapor viscosity, cP	1.196×10^{-2}	
Light liquid viscosity, cP	0.6	
Heavy liquid viscosity, cP	1.0	
Pressure, kPa	350	
Liquid Coalescer	No	
Required D, m	1.20	
Required TL, m	1.835	
Six inch spacing, m	0.1524	
LC length, m	6.694e−003	
LC Size, m	6.694e−003	

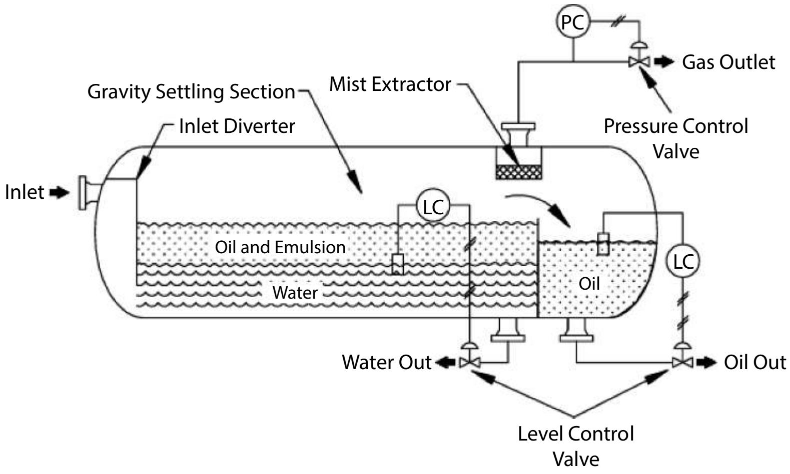


Figure 18.72 Schematic of a horizontal three-phase separator with interface level control and weir [79].

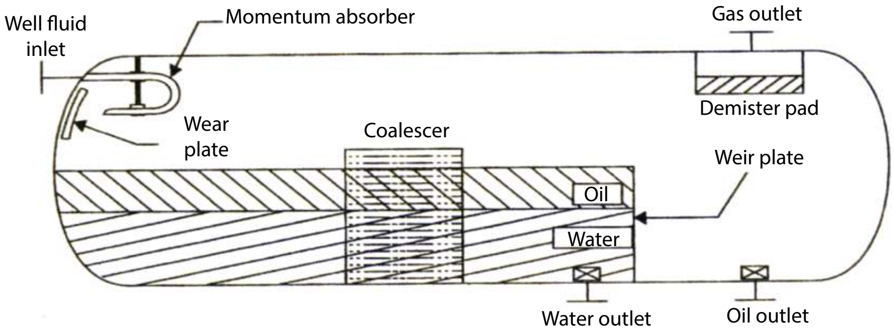


Figure 18.73 Three-phase separator [77].

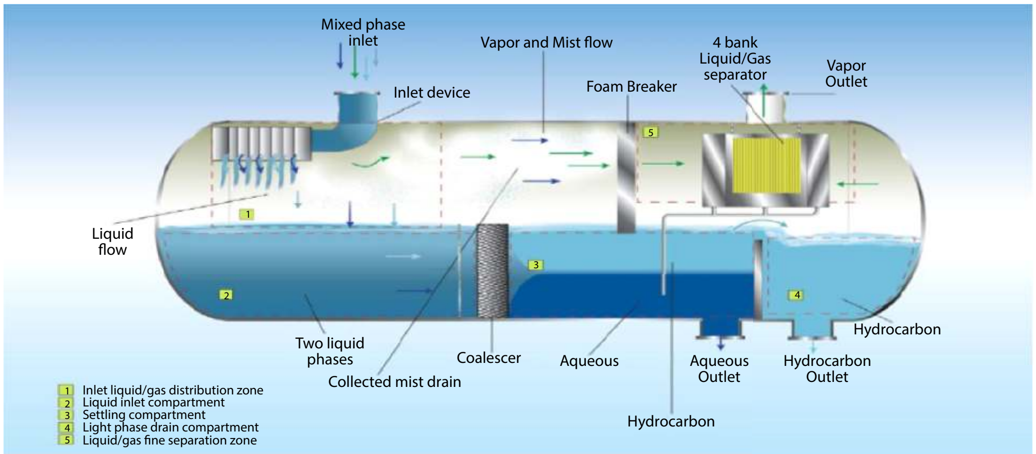


Figure 18.74 The configurations and notations of the horizontal three-phase separator (source: www.Petroleumbook.com 2017).

Separator Selection

API 12J recommends the following procedures for selecting a separator for a particular application [75]:

- Determine which shape fits the particular installation best, considering space, mounting and ease of access for maintenance. Both present and future operating conditions should be considered.
- Determine whether any unusual well stream conditions (foam, sand, and so on) would make the vessel selected difficult to operate or maintain.
- Check whether the cost of the vessel is reasonable.
- Check whether any heating coil is required against contamination by solid paraffin wax/ check hydrate formation.
- Check three-phase requirement for water removal.
- Check any possible liquid slugging of separator. Slugs are large lumps of liquid flowing along with gas.

Sizing Parameters and Guidelines

The three main functions considered in sizing three phase separators are as follows:

1. Provide sufficient residence time for the liquid so that degassing can occur.
2. Provide sufficient free cross-sectional area above the liquid so that the gas velocity is low enough to allow liquid particles to be separated.
3. Provide enough residence time for liquid–liquid separation to occur.

A separator sizing will be either liquid or gas controlling, which is determined by the gas/oil ratio. Water in oil can be quantified in terms of percentage and particle size. Although tight emulsions will cause problems, however a three-phase separator with a residence time of 3 min will produce an outlet oil quality of 3–8% water in oil. If the residence time is extended that results in an increase in price due to a larger vessel size, the oil quality can in most cases be improved. Other factors as heating prior to separation, addition of demulsifying agents, and so on, can also greatly assist with outlet oil quality improvement.

Gas in outlet oil does not usually raise any problem unless the gas is controlling due to a very high gas/oil ratio. Normally, a three-phase separator is liquid controlling in terms of sizing. Therefore, on letdown in pressure to the next stage of separation, liquids should not liberate any more gas than would theoretically exist under equilibrium conditions. The general requirement for sales of oil is 15 psia Reid vapor pressure, easily measured as per ASTM standard D323. The same parameters which dictate water in oil quality are relevant to oil in water quality. Additionally, if only a very small quantity of water is present in the formation, much tighter emulsions would tend to foam; with larger quantities, an interface can be readily established. As a guideline for settling time requirement the following table can be employed. Table 18.22 shows an interpolated chart for oil in water for various residence times.

Crude	**Outlet water in oil**
Heavy crude < 30°API	5000 ppm—for 3 min, 1000 ppm for 30 min
Medium crude 30°–45°API	3000 ppm—for 3 min, 1000 ppm for 10 min
Light crude > 45°API	1500 ppm—for 3 min

Table 18.22 Oil in water for various residence times.

Residence time (min)	**Density < 30°API**	**Density 37.5°API**	**Density 45°API**
3	5000 ppm	3000 ppm	1500 ppm
6	3000	1800	1000
10	1800	1080	500–1000
20	1080	500–1000	
30	500–1000		

Special considerations should be given to the sizing of separators under the following conditions:

Degassing of very light > 40°API dry oil.
Degassing and dehydration of waxy crude.
Handling of foaming crude.

Problems sometime occur when identifying crude that is likely to foam prior to specifying surface facilities. In general, crude oil is likely to foam if the crude specific gravity is less than 40°API, the temperature is less than 170°F and has a very low viscosity.

Separation Setup

There are two configurations of a separator, typically a horizontal and vertical separator. A horizontal separator is selected if there is large volume of total fluids and large quantities of dissolved gas predicted. Selecting a horizontal separator, fluid interface surface areas are maximized, allowing for entrapped gas to be released. This also provides the most efficient orientation for containing a large volume of incoming liquid flow.

The level settings within the vessel must be noted as these impact the criteria required for separation, i.e., residence time and surge time. The following level settings are required in a three-phase separator from a control and safety considerations:

LAHH	High liquid level trip
LAH	High liquid level alarm
NLL	Normal liquid level
LAL	Low liquid level alarm
LALL	Low liquid level trip
iLAH	High interface alarm
iNLL	Normal interface alarm
iLAL	Low interface alarm
iLALL	Low interface trip

Vessels can also be provided with an interface high level trip, but this is not always required and design codes should be consulted for clarification. Figure 18.75 shows level settings of a separator and Figure 18.76 shows a three-phase separator with level settings (submerged weir) [77].

High, Very High and Low, Very Low Levels for Instrumentation and Control

Figure 18.77 shows the positions of high and low levels to be fixed during the design stage. The tripping or emergency shutdown of the separator takes place where HLSD and LLSD values occur. Alarms occur at HLA and LLA values, sufficiently prior to the occurrence of tripping so that the operator can take the necessary corrective action. The table given in Figure 18.77 provides a rule of thumb guidance for the residence time in minutes between levels.

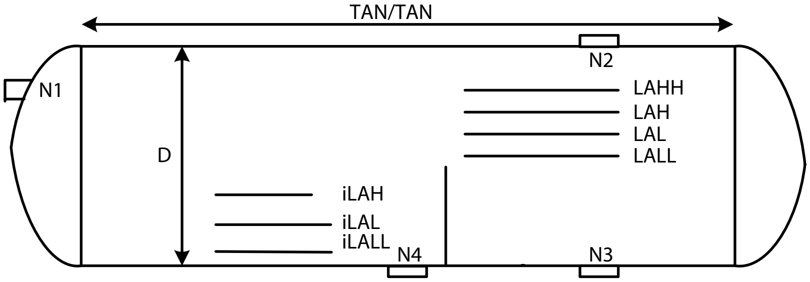
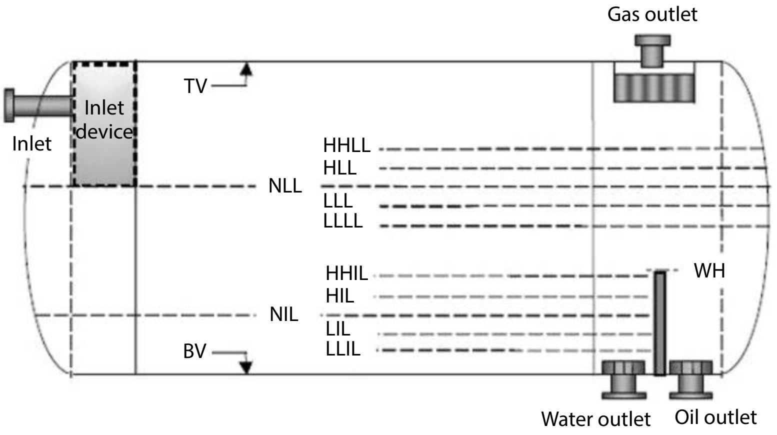
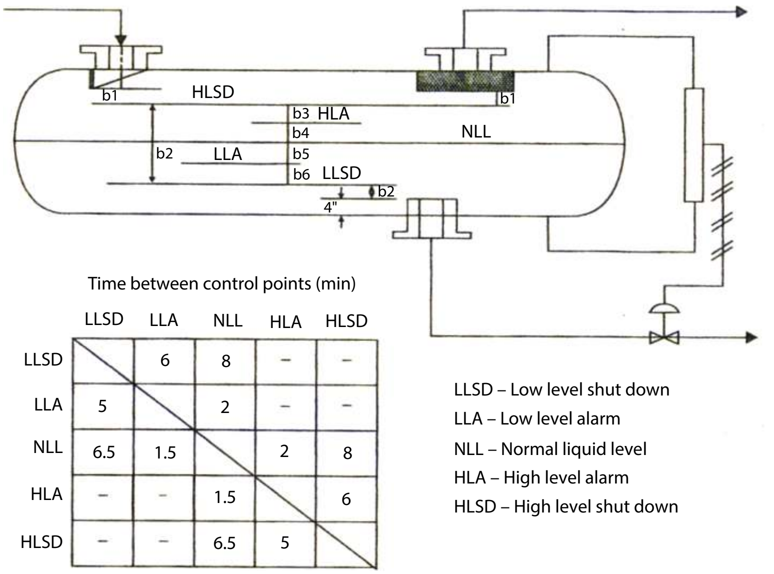


Figure 18.75 Separator level settings [84].



TV	Top of vessel
HHLL	High-high liquid level
HLL	High liquid level
NLL	Normal liquid
LLL	Low liquid level
LLLL	Low-low liquid level
WH	Weir height
HHIL	High-high interface level
HIL	High interface level
NIL	Normal interface level
LIL	Low interface level
LLIL	Low-low interface level
BV	Bottom of vessel
D_r	Inside diameter

Figure 18.76 Three-phase separator with level settings (submerged weir) [76].



Time between control points (min)

	LLSD	LLA	NLL	HLA	HLSD
LLSD		6	8	–	–
LLA	5		2	–	–
NLL	6.5	1.5		2	8
HLA	–	–	1.5		6
HLSD	–	–	6.5	5	

LLSD – Low level shut down
LLA – Low level alarm
NLL – Normal liquid level
HLA – High level alarm
HLSD – High level shut down

Figure 18.77 Guide for low-level and high-level alarms and trips [77].

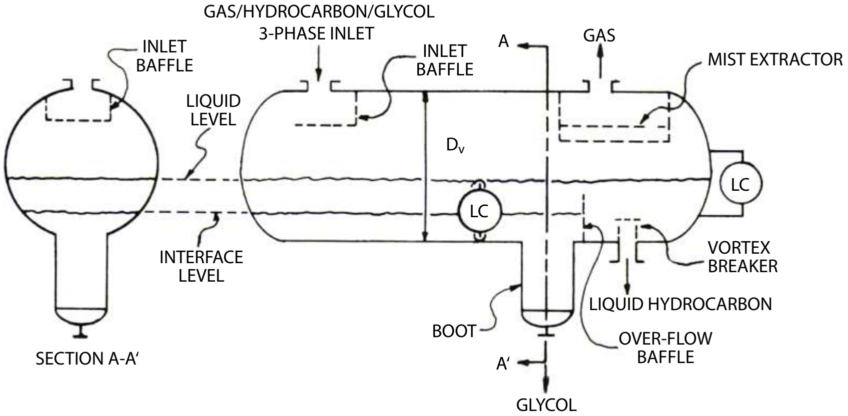


Figure 18.78 Horizontal three-phase separator with wire mesh mist extractor (source: Gas Processing Suppliers Association (GPSA) Engineering Databook, 12th ed. Section 7) [82].

Figure 18.79 Photograph of a three-phase separator [84].

The values to the right of the diagonal are for surging flow or feed to a fired heater. The values to the left are for normal operation [84]. Figure 18.78 shows a horizontal three-phase separator with wire mesh mist extractor. The figure illustrates the separation of two liquid phases (glycol and hydrocarbon). The denser glycol settles to the bottom and is withdrawn through the boot. The glycol level is controlled by an interface level control instrument. Figure 18.79 shows a photograph of a horizontal separator three-phase separator.

Sizing Three-Phase Oil–Gas Separator

Various combinations of vessel diameter and effective length are calculated based on gas capacity constraint and retention time constraints of oil and water. In the case of gas separation, the concepts pertaining to two-phase separation as illustrated earlier are valid. The maximum diameter of the separator is determined based on the need to settle typically 500-μm water droplets from the oil.

The retention (residence) time of oil and water provided determines the amount of water content in oil and oil content in produced water. The oil capacity provided should ensure that the oil reaches equilibrium with the flashed gas being liberated and the free water has time to coalesce into droplet sizes sufficient to settle. The water capacity provided should ensure that most of the large droplets of the oil entrained in water have sufficient time to coalesce and rise to the oil–water interface. An acceptable combination of diameter and length is chosen on the basis of the above constraints. A slenderness ratio between 3 and 5 is used for horizontal three-phase separators.

Figure 18.80a shows a typical configuration for a vertical three-phase separator. Flow enters the vessel through the side as in the horizontal separator. The inlet diverter separates the bulk of the gas. A downcomer is required to route

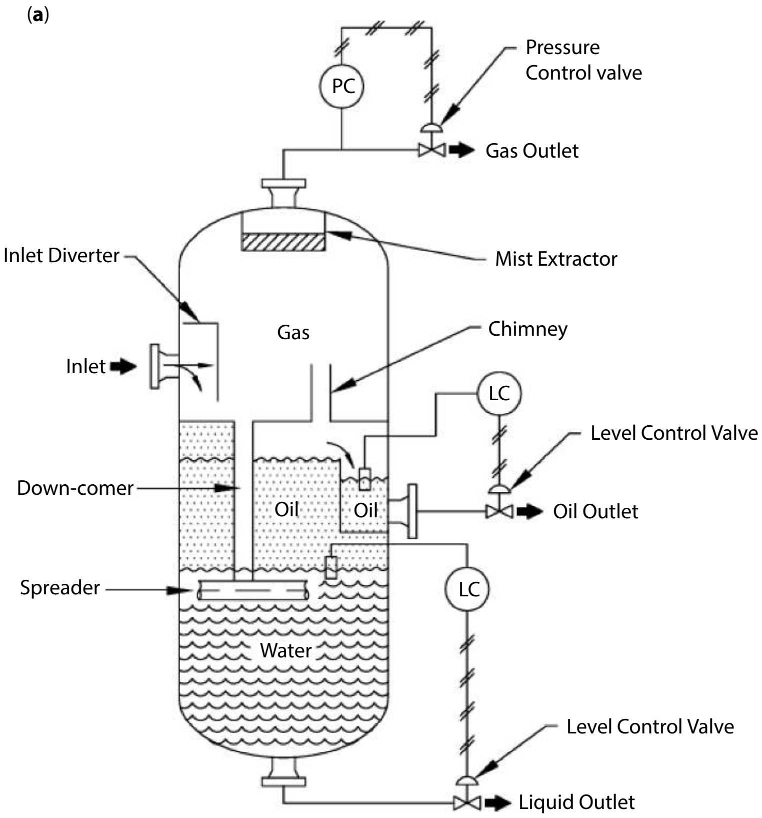


Figure 18.80a Schematic of a vertical three-phase separator with interface level control [79].

the liquid through the oil–gas interface so as not to disturb the oil skimming action taking place. A chimney is needed to equalize gas pressure between the lower section and the gas section.

The downcomer outlet is located just below the oil–water interface, thereby "water washing" the incoming stream. From this point as the oil rises, any free water trapped within the oil phase separates out. The water droplets flow countercurrent to the oil. The water flows downward and oil droplets trapped in the water phase tend to rise countercurrent to the water flow. Figures 18.80b and 18.80c are views of vertical three-phase separators without water washing and with interface control. Figure 18.80d shows the three different methods of control that are often used on vertical separators.

Procedure for Vertical Separator

As with vertical two-phase separators, a minimum diameter must be maintained to assure adequate gas capacity. Additionally, vertical three phase separators must maintain a minimum diameter to allow the 500 μm water droplets to settle. The height of the three-phase separator is determined from retention time considerations. The first step in sizing a vertical separator is to establish the design basis. This includes specifying the maximum and minimum flow rates, operating pressure and temperature, droplet size to be removed.

Gas Capacity Constraint

Setting the gas velocity equal to the terminal settling velocity of a droplet, the following equations are:

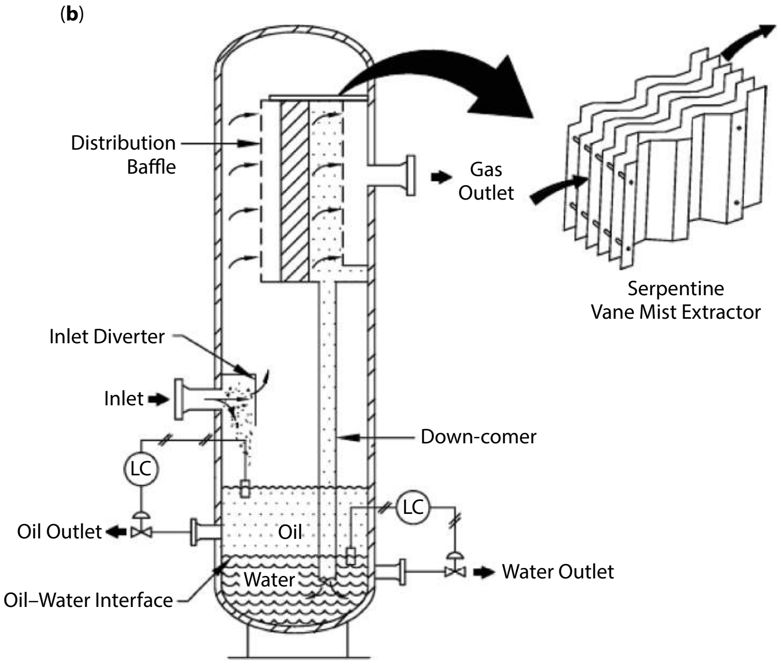


Figure 18.80b Cutaway view of the vertical three-phase separator without water washing and with vane mist extractor [79].

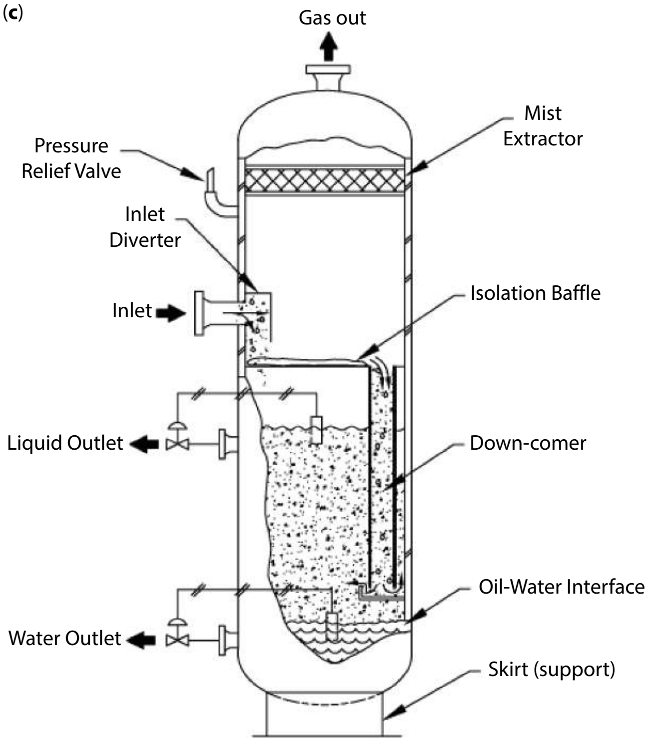


Figure 18.80c Cutaway view of the vertical three-phase separator without water washing and with wire-mesh mist extractor [79].

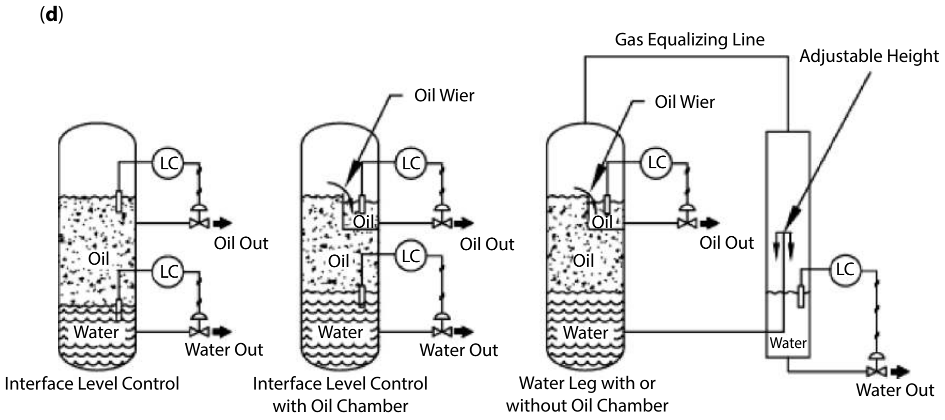


Figure 18.80d Liquid level control schemes [79].

Field Units

$$d^2 = 5{,}040\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l-\rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \tag{18.174}$$

SI Units

$$d^2 = 34{,}500\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l-\rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \tag{18.175}$$

From 100 μm droplet removal, Eqs. 18.174 and 18.175 are reduced to the following equations:

$$d^2 = 504\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l-\rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \tag{18.176}$$

SI Units

$$d^2 = 3{,}500\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l-\rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \tag{18.177}$$

Settling

The requirement for settling water droplets from the oil requires that the following equation must be satisfied:

Field Units

$$d^2 = \frac{6690\,Q_o\,\mu}{(\Delta SG)\,d_m^2} \tag{18.178}$$

SI Units

$$d^2 = 6.37 \times 10^8 \frac{Q_o \mu}{(\Delta SG) d_m^2} \tag{18.179}$$

Field Units

V_t is in ft/s, V_o in ft/s, d_m in micron, μ in cP

$$V_t = V_o$$

$$V_t = \frac{1.78 \times 10^{-6} (\Delta SG) d_m^2}{\mu} \tag{18.180}$$

Q is in ft^3/s, A in ft^2

$$V_o = \frac{Q}{A} \tag{18.181}$$

Q_o is in bpd

$$Q = Q_o \times 5.61 \frac{ft^3}{barrel} \times \frac{day}{24h} \times \frac{h}{3600s} = 6.49 \times 10^{-5} Q_o \tag{18.182}$$

D is in ft, d in inches.

$$A = \frac{\pi D^2}{4} = \frac{\pi d^2}{(4)(144)} \tag{18.183}$$

$$V_o = 0.0119 \frac{Q_o}{d^2} \tag{18.184}$$

Equating Eqs. 18.180 and 18.154 gives:

$$\frac{1.78 \times 10^{-6} (\Delta SG) d_m^2}{\mu} = 0.0119 \frac{Q_o}{d^2} \tag{18.185}$$

$$d^2 = \frac{6690 Q_o \mu}{(\Delta SG) d_m^2} \tag{18.178}$$

SI Units

V_t is in m/s, V_o in m/s, d_m in micron, μ in cP

$$V_t = V_o$$

$$V_t = \frac{5.556 \times 10^{-7} (\Delta SG) d_m^2}{\mu} \qquad (18.186)$$

Q is in m^3/s, A in m^2

$$V_o = \frac{Q}{A} \qquad (18.187)$$

Q_o is in (m^3/s)

$$Q = Q_o \times \frac{h}{3600s} = \frac{Q_o}{3600} \qquad (18.188)$$

D is in (m), d in (mm).

$$A = \frac{\pi D^2}{4} = \frac{\pi d^2}{(4)(1000)^2} \qquad (18.189)$$

$$V_o = 354 \frac{Q_o}{d^2} \qquad (18.190)$$

Equating Eqs. 18.186 and 18.190 gives:

$$\frac{5.556 \times 10^{-7} (\Delta SG) d_m^2}{\mu} = 353.6 \frac{Q_o}{d^2} \qquad (18.191)$$

$$d^2 = 6.37 \times 10^8 \frac{Q_o \mu}{(\Delta SG) d_m^2} \qquad (18.179)$$

For 500-μm droplets Eqs. 18.178 and 18.179 become

Field Units

$$d^2 = 0.0267 \left[\frac{Q_o \mu}{(\Delta SG)} \right] \qquad (18.192)$$

SI Units

$$d^2 = 2550 \left[\frac{Q_o \mu}{(\Delta SG)} \right] \qquad (18.193)$$

Settling Oil From Water Phase

The requirement for separating oil from water requires that the following equations must be satisfied.

Field Units

$$d^2 = \frac{6690 Q_o \mu}{(\Delta SG) d_m^2} \tag{18.178}$$

SI Units

$$d^2 = 6.37 \times 10^8 \frac{Q_o \mu}{(\Delta SG) d_m^2} \tag{18.179}$$

For 200-μm droplets, Eqs. 18.178 and 18.179 become:

Field Units

$$d^2 = 0.167 \left[\frac{Q_o \mu}{(\Delta SG)} \right] \tag{18.194}$$

SI Units

$$d^2 = 1.59 \times 10^4 \left[\frac{Q_o \mu}{(\Delta SG)} \right] \tag{18.195}$$

Retention Time Constraint

$$h_o + h_w = \frac{[(t_r)_o Q_o + (t_r)_w Q_w]}{0.12 d^2} \tag{18.196}$$

SI Units

$$h_o + h_w = \frac{[(t_r)_o Q_o + (t_r)_w Q_w]}{4.713 \times 10^{-8} d^2} \tag{18.197}$$

where

h_o = height of oil pad, in (mm)
h_w = height from water outlet to interface, in. (mm)
d = vessel's internal diameter, in. (mm)

Note: This height must be adjusted for cone bottom vessels.

From two-phase separator design:

Field Units

$$d^2h = \frac{t_r Q_l}{0.12} \tag{18.159}$$

Thus

$$d^2h_o = \frac{(t_r)_o Q_o}{0.12} \tag{18.198}$$

$$d^2h_w = \frac{(t_r)_w Q_w}{0.12} \tag{18.199}$$

Adding Eqs. 18.198 and 18.199 gives:

$$h_o + h_w = \frac{(t_r)_o Q_o + (t_r)_w Q_w}{0.12d^2} \tag{18.200}$$

SI Units

$$d^2h = \frac{t_r Q_l}{4.713 \times 10^{-8}} \tag{18.160}$$

$$d^2h_o = \frac{(t_r)_o Q_o}{4.713 \times 10^{-8}} \tag{18.201}$$

$$d^2h_w = \frac{(t_r)_w Q_w}{4.713 \times 10^{-8}} \tag{18.202}$$

Adding Eqs. 18.201 and 18.202 gives:

$$h_o + h_w = \frac{(t_r)_o Q_o + (t_r)_w Q_w}{4.713 \times 10^{-8} d^2} \tag{18.203}$$

Seam-to-Seam Length

The seam-to-seam length (L_{ss}) of vertical vessels may be estimated based on the diameter and liquid height. Allowance must be made for the gravity settling (gas separation) section, inlet diverter, mist extractor, and any space below the water outlet as shown in Figure 18.80e.

Field Units

$$L_{ss} = \frac{h_o + h_w + 76}{12} \text{ (for diameters} \leq 36 \text{ in.)} \tag{18.204}$$

SI Units

$$L_{ss} = \frac{h_o + h_w + 1930}{1000} \text{ (for diameters} \leq 914 \text{ mm)} \tag{18.205}$$

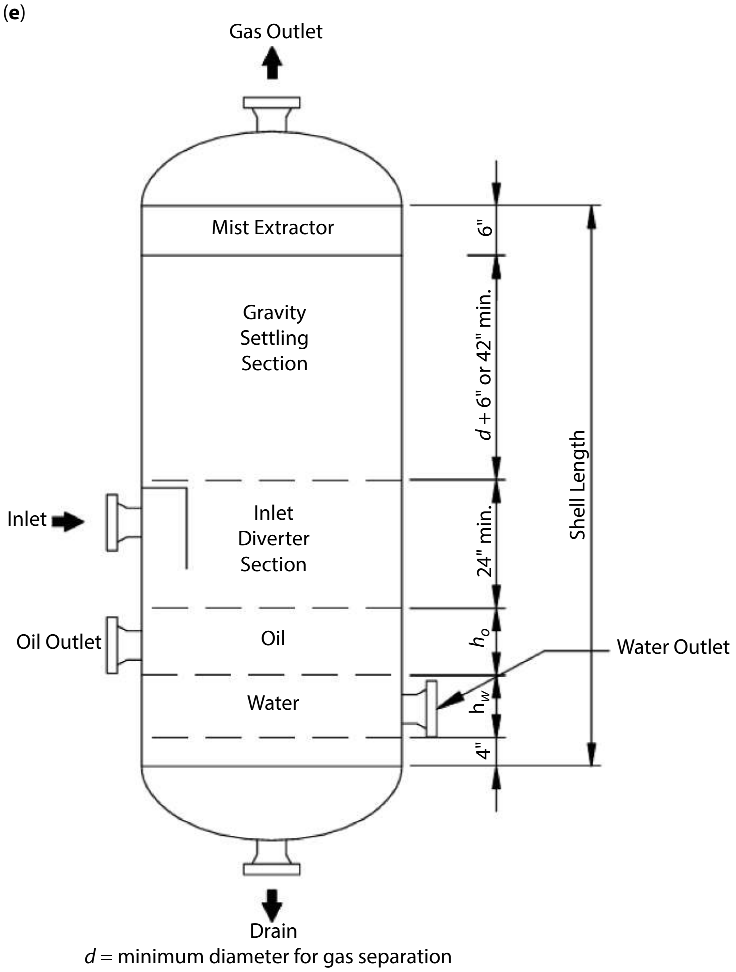


Figure 18.80e Approximate seam-to-seam shell length for a vertical three-phase separator [79].

Field Units

$$L_{ss} = \frac{h_o + h_w + d + 40}{12} \text{ (for diameters} > 36 \text{ in.)} \qquad (18.206)$$

SI Units

$$L_{ss} = \frac{h_o + h_w + d + 1016}{1000} \text{ (for diameters} > 914 \text{ mm)} \qquad (18.207)$$

where

h_o = height of oil pad, in (mm)
h_w = height from water outlet to interface, in. (mm)
d = vessel's internal diameter, in. (mm)

The larger of the L_{ss} values from Eqs. 18.204–18.207 should be used.

Slenderness Ratio

The larger the slenderness ratio, the less expensive the vessel. Sizing of vertical separators is dependent of liquid dominated and is common to choose slenderness ratios not greater than 4. In order to keep the height of the liquid collection section to a reasonable level, choices of 1.5–3.0 are common, although height restrictions may impose the choice of a lower slenderness ratio.

Procedure for Sizing Three-Phase Vertical Separators

1. Establish the design basis that includes specifying the minimum and maximum flow rates, operating pressure and temperature, droplet size to be removed and so on.
2. Eqs. 18.174 and 18.174 may be used to calculate the minimum diameter for a liquid droplet to fall through the gas phase. Use Eqs. 18.178 and 18.177 for 100-μm droplets if no other information is available.
3. Eqs. 18.178 and 18.179 can be used to calculate the minimum diameter for water droplets to fall through the oil phase. Eqs. 18.192 and 18.193 for 500-μm droplets if no other information is available.
4. Eqs. 18.178 and 18.179 can be used to calculate the minimum diameter for oil droplets to rise through the water phase. Eqs. 18.194 and 18.195 for 200-μm droplets if no other information is available.
5. Select the largest of the three diameters; calculate in steps 2–4 as the minimum diameter. Any value larger than this minimum may be used for the vessel diameter.
6. For the selected diameter and assumed values of $(t_r)_o$ and $(t_r)_w$, Eqs. 18.200 and 18.203 can be used to determine $h_o + h_w$.
7. From d and $h_o + h_w$, the seam-to-seam length can be estimated using Eqs. 18.204–18.206. The larger value of L_{ss} should be used.
8. Check the slenderness ratio. Slenderness ratios between 1.5 and 3 are common.
9. If possible, select a standard-size diameter and seam-to-seam length.

Example 18.14 Size a Vertical Three-Phase Separator (Field Units)

Given:

Q_o = 5000 bpd
Q_w = 3000 bpd
Q_g = 5 MMscfd
P = 100 psia
T = 90°F (550°R)
SG_{oil} = 30°API
SG_w = 1.07
S_g = 0.6
$(t_r)_o = (t_r)_w$ = 10 min
μ_o = 10 cP
μ_w = 1 cP
ρ_g = 0.298 lb/ft³
ρ_l = 54.67 lb/ft³
C_D = 2.01
Z = 0.99

Droplet removal = 100-μm liquids, 500-μm water, 200-μm oil.

Solution

1. Calculate the difference in specific gravities

$$^{\circ}\mathrm{API} = \frac{141.5}{(\mathrm{SG})_o} - 131.5$$

$$(\mathrm{SG})_o = \frac{141.5}{[^{\circ}\mathrm{API} + 131.5]} = \frac{141.5}{[30 + 131.5]} = 0.876$$

$$\Delta \mathrm{SG} = 1.07 - 0.876 = 0.194$$

2. Calculate the minimum diameter required to settle a liquid droplet through the gas phase from Eq. 18.174

$$d^2 = 5040\left[\frac{(550)(0.99)(5)}{(100)}\right]\left[\left(\frac{0.2975}{(54.67 - 0.2975)}\right)\frac{2.211}{100}\right]^{0.5}$$

$$d = 38.85 \text{ in.}$$

3. Calculate the minimum diameter required for water droplets to settle through the oil phase from Eq. 18.178.

$$d^2 = 6{,}690\left[\frac{(5{,}000)(10)}{(0.194)(500)^2}\right]$$

$$d = 83.0 \text{ in.}$$

Table 18.23 Vertical three-phase separator capacity diameter vs. length for retention time constraint $(t_r)_o = (t_r)_w$ = 10 min.

d_o (in.)	$h_o + h_w$ (in.)	L_{ss} (ft)	$S_R = \left(\frac{12L_{ss}}{d}\right)$
84	94.5	18.2	2.6
90	82.3	17.7	2.4
96	72.3	17.4	2.2
102	64.1	17.2	2

4. Calculate the minimum diameter required for oil droplets to rise through the water phase from Eq. 18.178.

$$d^2 = 6{,}690\left[\frac{(3{,}000)(1)}{(0.194)(200)^2}\right]$$
$$d = 50.8 \text{ in.}$$

5. Select the largest diameter from steps 2–4 as the minimum inside diameter required

$$d_{min} = 83.0 \text{ in.}$$

6. Calculate $h_o + h_w$ from Eq. 18.200. Table 18.23 shows the results.

$$h_o + h_w = \frac{(10)[5{,}000 + 3{,}000]}{0.12d^2} = \frac{667{,}000}{d^2}$$

7. Calculate seam-to-seam length (L_{ss}) from Eq. 18.204 or Eq. 18.206. Since d > 36 in., use Eq. 18.204.

$$L_{ss} = \frac{h_o + h_w + d + 40}{12}$$

8. Calculate the slenderness ratio, S_R from Eq. 18.138 (Table 18.23). Choices in the range of 1.5–3 are common (Table 18.23).
9. Make a final selection: Calculate the combinations of d and $h_o + h_w$ for diameters greater than the minimum diameter (d_{min} = 83 in.). Table 18.23 shows the results. Select 90 in. outside diameter (OD) × 20 ft seam-to-seam length (L_{ss}).

The Excel spreadsheet Example 18.14.xlsx shows the calculations of Example 18.14, and the table below shows the results of the spreadsheet calculations.

Sizing Calculation of a vertical three-phase separator (Field Units)		
Oil flow rate	5000	bpd
Water flow rate	3000	bpd
Gas flow rate	5	MMscfd
SG of oil	30	°API
SG of water	1.07	
SG of gas	0.6	
Operating temperature	90	°F
Operating pressure	100	psia
Droplet size removal	140	micron
Retention time, $(t_r)_o = (t_r)_w$	10	min
Compressibility factor, Z	0.99	
Viscosity of oil, μ_o	10	cP
Viscosity of water, μ_w	1	cP
Assume Drag coefficient, C_D	2.01	
Droplet size removal in liquid	100	micron
Droplet size removal in water	500	micron
Droplet size removal in oil	200	micron
Viscosity, μ	0.013	cP
The terminal velocity, V_t	1.3009	ft/s
Reynolds number, Re:	20	
Drag coefficient, C_D:	2.2108	
Minimum diameter, d_{min}	83	in.
$h_o + h_w$	96.77	in.
Seam-to-seam length of the vesssel	18.31	ft

Example 18.15 Size a Vertical Three-Phase Separator (SI Units)

Given:

Q_o = 33 m^3/h
Q_w = 19.8 m^3/h
Q_g = 5902 sm^3/h
P = 6900 kPa
T = 32.2°C (305.4K)
SG_{oil} = 30°API
SG_w = 1.07
S_g = 0.6
$(t_r)_o = (t_r)_w$ = 10 min

μ_o = 10 cP
μ_w = 1 cP
ρ_g = 4.782 kg/m^3
ρ_l = 876.16 kg/m^3
C_D = 2.211
Z = 0.99

Droplet removal = 100-μm liquids, 500-μm water, 200-μm oil.

Solution

1. Calculate the difference in specific gravities

$$^{o}\text{API} = \frac{141.5}{(\text{SG})_o} - 131.5$$

$$(\text{SG})_o = \frac{141.5}{[^{o}\text{API} + 131.5]} = \frac{141.5}{[30 + 131.5]} = 0.876$$

$$\Delta\text{SG} = 1.07 - 0.876 = 0.194$$

2. Calculate the minimum diameter required to settle a liquid droplet through the gas phase from Eq. 18.175

$$d^2 = 34{,}500\left[\frac{(305.35)(0.99)(5902)}{(690)}\right]\left[\left(\frac{4.782}{(876.16 - 4.782)}\right)\frac{2.211}{100}\right]^{0.5}$$

$$d = 991.3\ \text{mm}$$

3. Calculate the minimum diameter required for water droplets to settle through the oil phase from Eq. 18.179.

$$d^2 = 6.37 \times 10^8\left[\frac{(33)(10)}{(0.194)(500)^2}\right]$$

$$d = 2082\ \text{mm}$$

4. Calculate the minimum diameter required for oil droplets to rise through the water phase from Eq. 18.179.

$$d^2 = 6.37 \times 10^8\left[\frac{(19.8)(1)}{(0.194)(200)^2}\right]$$

$$d = 1275\ \text{mm.}$$

5. Select the largest diameter from steps 2–4 as the minimum inside diameter required

$$d_{min} = 2082\ \text{mm}$$

6. Calculate $h_o + h_w$ from Eq. 18.197. Table 18.24 shows the results.

Table 18.24 Vertical three-phase separator capacity diameter vs. length for retention time constraint $(t_r)_o = (t_r)_w = 10$ min.

d_o (mm)	$h_o + h_w$ (mm)	L_{ss} (m)	$S_R = \left(\frac{1000\,L_{ss}}{d}\right)$
2286	2144	5.4	2.4
2438	1884	5.3	2.2
2591	1668	5.3	2.0

$$h_o + h_w = \frac{(10)[33+19.8]}{4.713\times10^{-8}\,d^2} = \frac{1.12\times10^{10}}{d^2}$$

7. Calculate seam-to-seam length (L_{ss}) from Eq. 18.205 or Eq. 18.207. Since d > 914 mm, use Eq. 18.207.

$$L_{ss} = \frac{h_o + h_w + d + 1016}{1000}$$

8. Calculate the slenderness ratio, S_R from Eq. 18.139 (Table 18.24). Choices in the range of 1.5–3 are common (Table 18.24).
9. Make a final selection: Calculate the combinations of d and $h_o + h_w$ for diameters greater than the minimum diameter ($d_{min} = 2082$ mm). Table 18.24 shows the results. Select 2286 mm outside diameter (OD) × 5.4 m seam-to-seam length (L_{ss}).

The Excel spreadsheet Example 18.15.xlsx shows the calculations of Example 18.15, and the table below shows the results of the spreadsheet calculations.

Sizing Calculation of a vertical three-phase separator (S.I. Units)		
Oil flow rate	33	m^3/h
Water flow rate	19.8	m^3/h
Gas flow rate	5902	sm^3/h
SG of oil	30	°API
SG of water	1.07	
SG of gas	0.6	
Operating temperature	32.2	°C
Operating pressure	690	kPa
Droplet size removal	140	micron
Retention time, $(t_r)_o = (t_r)_w$	10	min
Compressibility factor, Z	0.99	
Viscosity of oil, μ_o	10	cP
Viscosity of water, μ_w	1	cP

Assume Drag coefficient, C_D	2.01	
Droplet size removal in liquid	100	micron
Droplet size removal in water	500	micron
Droplet size removal in oil	200	micron
Viscosity, μ	0.013	cP
The terminal velocity, V_t	0.41	m/s
Reynolds number, Re:	20	
Drag coefficient, C_D:	2.2108	
Minimum diameter, d_{min}	2081.9	mm
h_o+h_w	2584.74	mm
Seam-to-seam length of the vesssel	5.68	m

Horizontal Separator Sizing—Half Full

Sizing a horizontal three-phase separator requires specifying the vessel diameter and a seam-to-seam vessel length. The gas capacity and retention time considerations establish certain acceptable combinations of diameter and length. The maximum diameter corresponding to the given liquid retention time requires the need to settle 500-μm water droplets from the oil and 200-μm oil droplets from the water.

Gas Capacity Constraint

Settling the gas retention time equal to the time required for a drop to settle to the liquid interface requires the following equations.

Gas Capacity

The gas capacity constraints provide the following equation:

Field Units

$$dL_{eff} = 420\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l-\rho_g}\right)\frac{C_d}{d_m}\right]^{1/2} \tag{18.208}$$

SI Units

$$dL_{eff} = 34.5\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l-\rho_g}\right)\frac{C_d}{d_m}\right]^{1/2} \tag{18.209}$$

where

d, in. (mm) = vessel inside diameter
L_{eff}, ft. (m) = vessel effective length
T, °R (K) = operating temperature

Z = gas compressibility factor
Q_g, MMscfd (scm/h) = gas flow rate
P, psia (kPa) = operating pressure
ρ_g, lb_m/ft^3 (kgm^3) = gas density
ρ_l, lb_m/ft^3(kgm^3) = liquid density
C_D = drag coefficient
d_m, microns (μm) = liquid drop to be separated

Retention Time

Liquid retention time constraints can be used to develop the following equations that determine acceptable combinations of d and L_{eff}.

Field Units

$$d^2L_{eff} = 1.42[(t_r)_o\, Q_o + (t_r)_w\, Q_w] \tag{18.210}$$

SI Units

$$d^2L_{eff} = 4.2 \times 10^4\, [(t_r)_o\, Q_o + (t_r)_w\, Q_w] \tag{18.211}$$

where

Q_w, bpd (m^3/h) = water flow rate
$(t_r)_w$, min. = water retention time
Q_o, bpd (m^3/h) = oil flow rate
$(t_r)_o$, min = oil retention time

Deriving Eqs. 18.210 and 18.211 is as follows:

t in s, Vol in ft^3 (m^3), Q in ft^3/s (m^3/s), D in ft (m), d in in. (mm), L_{eff} in ft (m).

$$\begin{gathered} t = \frac{Vol}{Q} \\ Vol = \frac{1}{2}\left(\frac{\pi D^2 L_{eff}}{4}\right) = \frac{\pi d^2 L_{eff}}{(2)(4)(144)} = 2.73 \times 10^{-3}\, d^2 L_{eff} \end{gathered} \tag{18.212}$$

$$(Vol)_o = 2.73 \times 10^{-3}\, d^2 L_{eff}\left(\frac{A_o}{A_l}\right) \tag{18.213}$$

$$(Vol)_w = 2.73 \times 10^{-3}\, d^2 L_{eff}\left(\frac{A_w}{A_l}\right) \tag{18.214}$$

Q_o and Q_w are bpd

$$Q = Q_o \times 5.61 \frac{ft^3}{barrel} \times \frac{day}{24\,\cancel{h}} \times \frac{\cancel{h}}{3600s} \tag{18.215}$$

For oil:

$$Q = 6.49 \times 10^{-5}\, Q_o \tag{18.216}$$

$$t_o = \frac{(\mathrm{Vol})_o}{Q_o} = 42\frac{d^2 L_{eff}}{Q_o}\left(\frac{A_o}{A_l}\right) \tag{18.217}$$

For Water:

$$Q = 6.49 \times 10^{-5}\, Q_w \tag{18.218}$$

$$t_w = \frac{(\mathrm{Vol})_w}{Q_w} = 42\frac{d^2 L_{eff}}{Q_w}\left(\frac{A_w}{A_l}\right) \tag{18.219}$$

SI Units

$$\begin{gathered} t = \frac{\mathrm{Vol}}{Q} \\ \mathrm{Vol} = \frac{1}{2}\left(\frac{\pi D^2 L_{eff}}{4}\right) = \frac{\pi d^2 L_{eff}}{(2)(4)(1000^2)} = 3.927 \times 10^{-7}\, d^2 L_{eff} \end{gathered} \tag{18.220}$$

$$(\mathrm{Vol})_o = 3.927 \times 10^{-7}\, d^2 L_{eff}\left(\frac{A_o}{A_l}\right) \tag{18.221}$$

$$(\mathrm{Vol})_w = 3.927 \times 10^{-7}\, d^2 L_{eff}\left(\frac{A_w}{A_l}\right) \tag{18.222}$$

Q_o and Q_w are m^3/h

For Oil:

$$Q = Q_o \times \frac{h}{3600s} = \frac{Q_o}{3600} \tag{18.223}$$

$$t_o = \frac{(\mathrm{Vol})_o}{Q_o} = 0.0014\frac{d^2 L_{eff}}{Q_o}\left(\frac{A_o}{A_l}\right) \tag{18.224}$$

For water:

$$Q = \frac{Q_w}{3600} \tag{18.225}$$

$$t_w = \frac{(\mathrm{Vol})_w}{Q_w} = 0.0014\frac{d^2 L_{eff}}{Q_w}\left(\frac{A_o}{A_l}\right) \tag{18.226}$$

A_o, A_w, and A_l are cross-sectional areas of oil, water and liquid respectively.

Rearranging the equation for t_o and t_w

$$42\left(\frac{A_o}{A_l}\right)=\frac{t_o Q_o}{d^2 L_{eff}}, \qquad 42\left(\frac{A_w}{A_l}\right)=\frac{t_w Q_w}{d^2 L_{eff}}, \tag{18.227}$$

$(t_r)_o$ and $(t_r)_w$ and in min,

$$0.7\left(\frac{A_o}{A_l}\right)=\frac{(t_r)_o Q_o}{d^2 L_{eff}}, \qquad 0.7\left(\frac{A_w}{A_l}\right)=\frac{(t_r)_w Q_w}{d^2 L_{eff}} \tag{18.228}$$

Adding Eqs. 18.227 and 18.228 gives:

$$0.7\left(\frac{A_o+A_w}{A_l}\right)=\frac{[(t_r)_o Q_o+(t_r)_w Q_w]}{d^2 L_{eff}} \tag{18.229}$$

$$A_o + A_w = A_l \tag{18.230}$$

Substituting Eq. 18.230 into Eq. 18.229 and rearranging gives:

$$d^2L_{eff} = 1.42[(t_r)_o\, Q_o + (t_r)_w\, Q_w] \tag{18.231}$$

SI Units

Rearranging the equation for t_o and t_w

$$0.0014\left(\frac{A_o}{A_l}\right)=\frac{t_o Q_o}{d^2 L_{eff}}, \qquad 0.0014\left(\frac{A_w}{A_l}\right)=\frac{t_w Q_w}{d^2 L_{eff}}, \tag{18.232}$$

$(t_r)_o$ and $(t_r)_w$ and in min.

$$2.356\times10^{-5}\left(\frac{A_o}{A_l}\right)=\frac{(t_r)_o Q_o}{d^2 L_{eff}}, \qquad 2.356\times10^{-5}\left(\frac{A_w}{A_l}\right)=\frac{(t_r)_w Q_w}{d^2 L_{eff}} \tag{18.233}$$

Adding Eqs. 18.232 and 18.233 gives:

$$2.356\times10^{-5}\left(\frac{A_o+A_w}{A_l}\right)=\frac{[(t_r)_o Q_o+(t_r)_w Q_w]}{d^2 L_{eff}} \tag{18.234}$$

$$A_o + A_w = A_l \tag{18.235}$$

Substituting Eq. 18.235 into Eq. 18.234 and rearranging gives:

$$d^2L_{eff} = 4.2 \times 10^4 \, [(t_r)_o \, Q_o + (t_r)_w \, Q_w] \tag{18.236}$$

Settling Water Droplets From Oil Phase

The velocity of water droplets settling through oil can be determined by Stokes' law. From this velocity and the specified oil phase retention time, the distance that a water droplet can settle may be estimated. This settling distance establishes a maximum oil pad thickness given by the following:

Field Units

$$h_o = \frac{0.00128(t_r)_o(\Delta SG)d_m^2}{\mu} \tag{18.237}$$

SI Units

$$h_o = \frac{0.033(t_r)_o(\Delta SG)d_m^2}{\mu} \tag{18.238}$$

t_w, t_o are in s, V is ft/s (m/s), h_o is in. (mm), d_m is μm, μ is cP

$$t_w = t_o$$

Field Units

$$t_w = \frac{h_o/12}{V_t}, \quad V_t = \frac{1.78 \times 10^{-6}(\Delta SG)d_m^2}{\mu} \tag{18.239}$$

$$t_w = \frac{\mu \, h_o}{(12)(1.78 \times 10^{-6})(\Delta SG)d_m^2} \tag{18.240}$$

$$t_w = 46816.47 \frac{\mu \, h_o}{(\Delta SG)d_m^2} \tag{18.241}$$

t_r is min

$$t_o = 60 \, (t_r)_o \tag{18.242}$$

Equating Eqs. 18.241 and 18.242 gives:

$$46816.47 \frac{\mu \, h_o}{(\Delta SG)d_m^2} = 60(t_r)_o \tag{18.243}$$

Rearranging Eq. 18.243 gives:

$$h_o = \frac{1.28 \times 10^{-3} (t_r)_o (\Delta SG) d_m^2}{\mu} \tag{18.244}$$

SI Units

$$t_w = \frac{h_o/1000}{V_t}, \qquad V_t = \frac{5.556 \times 10^{-7} (\Delta SG) d_m^2}{\mu} \tag{18.245}$$

$$t_w = 1800 \frac{\mu h_o}{(\Delta SG) d_m^2} \tag{18.246}$$

t_r in min.

$$t_o = 60\ (t_r)_o \tag{18.247}$$

Equating Eqs. 18.246 and 18.247 gives:

$$1800 \frac{\mu h_o}{(\Delta SG) d_m^2} = 60 (t_r)_o \tag{18.248}$$

Rearranging Eq. 18.248 gives;

$$h_o = \frac{0.033 (t_r)_o (\Delta SG) d_m^2}{\mu} \tag{18.249}$$

This is the maximum thickness the oil pad can be and still allows the water droplets to settle out in time $(t_r)_o$. For d_m = 500-μm, the following equation may be used.

Field Units

$$(h_o)_{max} = 320 \frac{(t_r)_o (\Delta SG)}{\mu} \tag{18.250}$$

SI Units

$$(h_o)_{max} = 8250 \frac{(t_r)_o (\Delta SG)}{\mu} \tag{18.251}$$

For a given oil retention time $(t_r)_o$ and a given water retention time $(t_r)_w$, the maximum oil pad thickness constraint establishes a maximum diameter in accordance with the following procedure:

1. Calculate $(h_o)_{max}$. Use 500-μm droplet if no other information is available.
2. Calculate the fraction of the vessel cross-sectional area occupied by the water phase. This given by

$$\frac{A_w}{A} = \frac{0.5[Q_w(t_r)_w]}{[(t_r)_o Q_o + (t_r)_w Q_w]} \tag{18.252}$$

3. From Figure 18.81, determine the coefficient β.
4. Calculate d_{max} from

$$d_{max} = \frac{(h_o)_{max}}{\beta} \tag{18.253}$$

where

$$\beta = \frac{(h_o)}{d} \tag{18.254}$$

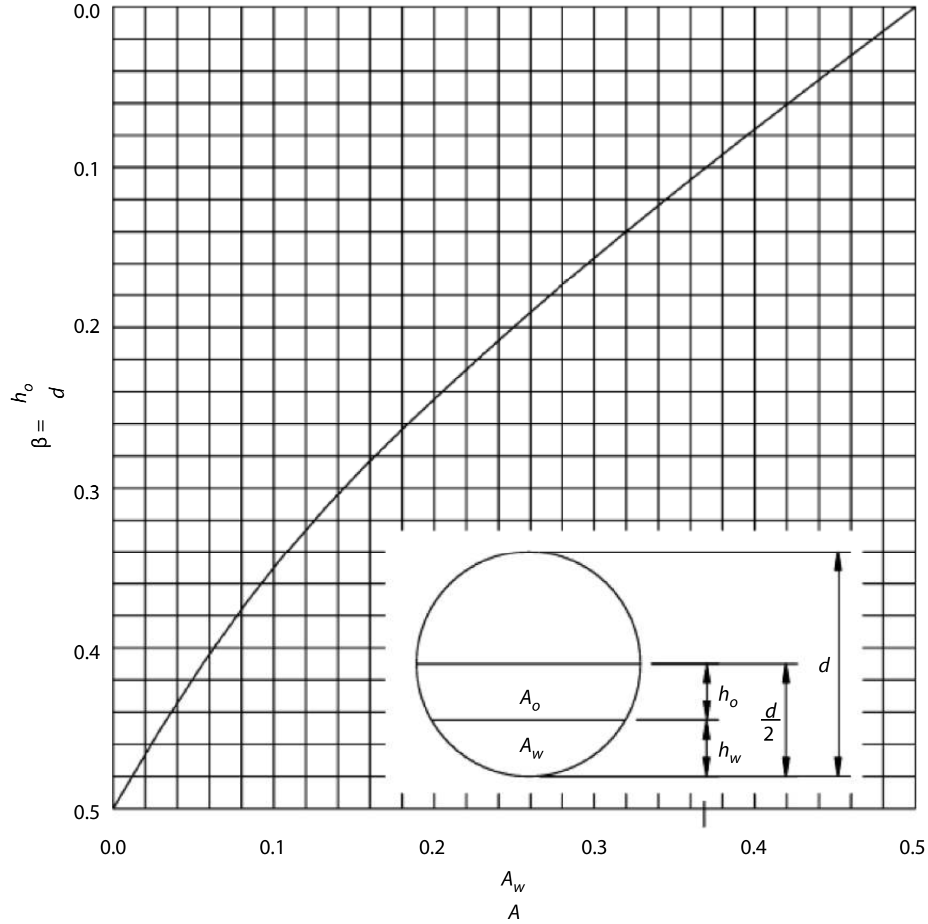


Figure 18.81 Coefficient “β” for a cylinder half filled with liquid [79].

Any combination of d and L_{eff} that satisfies Eqs. 18.208–18.211 and Eqs. 18.237–18.238 will meet the necessary criteria.

Derivation of Eq. 18.252 is as follows:

Field Units

A_o and A_w are in ft^2, Q in ft^3/s, t in s. and L_{eff} in ft.

$$A = \frac{Q \cdot t}{L_{eff}} \tag{18.255}$$

$$Q = 6.49 \times 10^{-5}\, Q_o, \quad Q = 6.49 \times 10^{-5}\, Q_w \tag{18.256}$$

$$t_o = 60(t_r)_o, \quad t_w = 60(t_r)_w \tag{18.257}$$

$$A_o = 3.89 \times 10^{-3} \frac{Q_o (t_r)_o}{L_{eff}}, \quad A_w = 3.89 \times 10^{-3} \frac{Q_w (t_r)_w}{L_{eff}} \tag{18.258}$$

For a vessel ½ full of liquid:

$$A = 2(A_o + A_w) \tag{18.259}$$

$$\frac{A_w}{A} = \frac{0.5[Q_w (t_r)_w]}{[(t_r)_o\, Q_o + (t_r)_w\, Q_w]} \tag{18.252}$$

SI Units

A_o and A_w are in m^2, Q in m^3/s, t in s and L_{eff} in m.

$$A = \frac{Q \cdot t}{L_{eff}} \tag{18.255}$$

$$Q = \frac{Q_o}{3600}, \quad Q = \frac{Q_w}{3600} \tag{18.260}$$

$$t_o = 60(t_r)_o, \quad t_w = 60(t_r)_w \tag{18.257}$$

$$A_o = 0.0167 \frac{Q_o (t_r)_o}{L_{eff}}, \quad A_w = 0.0167 \frac{Q_w (t_r)_w}{L_{eff}} \tag{18.261}$$

For a vessel ½ full of liquid:

$$A = 2(A_o + A_w) \tag{18.259}$$

$$\frac{A_w}{A} = \frac{0.5[Q_w (t_r)_w]}{[(t_r)_o\, Q_o + (t_r)_w\, Q_w]} \tag{18.252}$$

Separating Oil Droplets from Water Phase

Oil droplets in the water phase rise at a terminal velocity defined by Stokes' law. As with water droplets in oil, the vessel and retention time may be used to determine a maximum vessel diameter from Eqs. (18.210) and (18.211). Generally, the maximum diameter determined from a 200-μm oil droplet rising through the water phase is seldom larger than 500-μm water droplet falling through the oil phase. Therefore, the maximum diameter determined from a 500-μm water droplet settling through the oil phase normally governs the vessel design. For $d_m = 200$ μm, the following equations may be used:

Field Units

$$(h_w)_{max} = \frac{51.2(t_r)_w\,(\Delta SG)}{\mu_w} \tag{18.262}$$

SI Units

$$(h_w)_{max} = \frac{1{,}520(t_r)_w\,(\Delta SG)}{\mu_w} \tag{18.263}$$

The maximum diameter may be found from the following equation:

$$d_{max} = \frac{(h_w)_{max}}{\beta} \tag{18.264}$$

Note: d_{max} depends on Q_o, Q_w, $(t_r)_o$, and $(t_r)_w$

Seam-to-Seam Length

The effective length may be calculated from Eqs. 18.210 and 18.211, and a vessel seam-to-seam length may then be estimated. The actual required seam-to-seam length is dependent on the physical design of the vessel. For vessels sized based on gas capacity, some portion of the vessel length is required to distribute the flow evenly near the inlet diverter. Another portion of the vessel length is required for the mist extractor. The length of the vessel between the inlet and the mist extractor with evenly distributed flow is the L_{eff} calculated from Eqs. 18.208 and 18.209. As the vessel's diameter increases, more length is required to evenly distribute the gas flow. Independent on how small the diameter may be, a portion of length is still required for the both the flow distribution and the mist extractor. Figure 18.82 shows an approximate seam-to-seam shell length of a horizontal separator one-half full [79].

The seam-to-seam length of a vessel may be estimated as the larger of the

$$L_{ss} = \frac{4}{3} L_{eff}\,(\text{for liquid capacity}) \tag{18.265}$$

Field Units

$$L_{ss} = L_{eff} + \frac{d}{12}(\text{for gas capacity}) \tag{18.266}$$

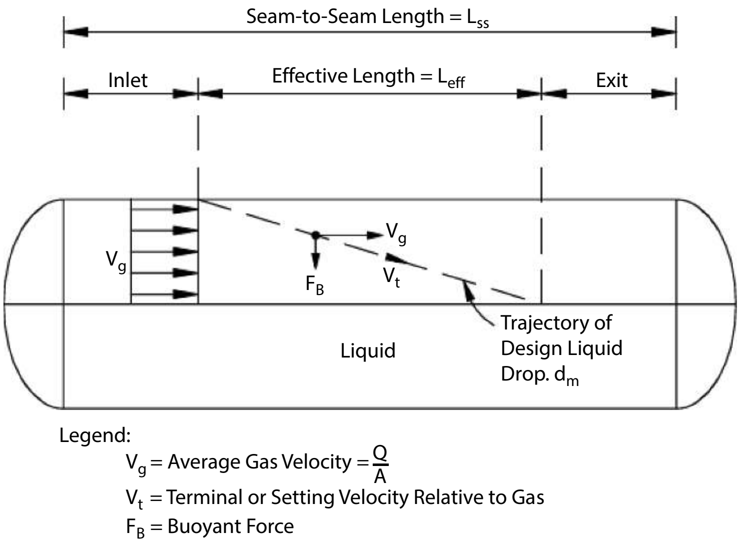


Figure 18.82 Approximate seam-to-seam shell length of a horizontal separator one-half full [79].

SI Units

$$L_{ss} = L_{eff} + \frac{d}{1000}(\text{for gas capacity}) \tag{18.267}$$

For vessels sized on a liquid capacity basis, some portion of the vessel length is required for inlet diverter flow distribution and liquid outlet. The seam-to-seam length should not exceed the following:

$$L_{ss} = \frac{4}{3}L_{eff} \tag{18.265}$$

Slenderness Ratio

For each vessel design, a combination of L_{eff} and d exists that will minimize the cost of the vessel. Generally, the smaller the diameter of a vessel, the less it will cost. Decreasing the diameter increases the fluid velocities and turbulence. As the vessel's diameter decreases, there is the possible gas re-entraining liquids or destruction of the oil/water interface increases. The ratio of the seam-to-seam length divided by the outside diameter is referred to as the "slenderness ratio" (S_R) of the vessel and should be between 3 and 5. If the ratio is outside the 3 to 5 range is used; the design should be checked to ensure that re-entrainment does not occur.

Procedure for Sizing Three-Phase Horizontal Separators—Half-Full

1. Establish the design basis, which includes specifying the maximum and minimum flow rates, operating pressure and temperature, droplet size to be removed and so on.
2. Select a $(t_r)_o$ and a $(t_r)_w$.
3. Calculate $(h_o)_{max}$. Use a 500-μm droplet if no other information is available.

Field Units

$$(h_o)_{max} = 1.28\times10^{-3}\frac{(t_r)_o(\Delta SG)d_m^2}{\mu}$$

For 500-μm

$$(h_o)_{max} = 320\frac{(t_r)_o(\Delta SG)}{\mu}$$

SI Units

$$(h_o)_{max} = 0.033\frac{(t_r)_o(\Delta SG)d_m^2}{\mu}$$

For 500-μm

$$(h_o)_{max} = 8250\frac{(t_r)_o(\Delta SG)}{\mu}$$

4. **4.** Calculate A_w/A:
5. **5.** Determine β from curve.
6. **6.** Calculate d_{max}:

$$d_{max} = \frac{(h_o)_{max}}{\beta}$$

Note: d_{max} depends on Q_o, Q_w, $(t_r)_o$, and $(t_r)_w$.

7. **7.** Calculate combinations of d, L_{eff} for d less than d_{max} that satisfy the gas capacity constraint. Use 100-μm droplet if no other information is available.

Field Units

$$dL_{eff} = 420\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l-\rho_g}\right)\frac{C_d}{d_m}\right]^{1/2} \tag{18.208}$$

SI Units

$$dL_{eff} = 34.5\left[\frac{TZQ_g}{P}\right]\left[\left(\frac{\rho_g}{\rho_l-\rho_g}\right)\frac{C_d}{d_m}\right]^{1/2} \tag{18.209}$$

8. **8.** Calculate combinations of d, L_{eff} for d less than d_{max} that satisfy the oil and water retention time constraints.

9. Estimate seam-to-seam length.

Field Units

$$L_{ss} = \frac{4}{3} L_{eff} \text{(for liquid capacity)} \tag{18.265}$$

$$L_{ss} = L_{eff} + \frac{d}{12} \text{(for gas capacity)} \tag{18.266}$$

SI Units

$$L_{ss} = \frac{4}{3} L_{eff} \text{(for liquid capacity)} \tag{18.265}$$

$$L_{ss} = L_{eff} + \frac{d}{1000} \text{(for gas capacity)} \tag{18.267}$$

10. Select a reasonable diameter and length. Slenderness ratios (12 L_{ss}/d) on the order of 3.5 are common.
11. Making a final selection, always use a standard vessel size. API Spec. 12J uses small separators. In larger sizes, heads come in outside diameters, which are multiplies of 6 in. (150 mm). The width of steel sheets for the shells is usually 10 ft (3000 mm), therefore it is common practice sot specify L_{ss} in multiples of 5 [75].

Horizontal Separators Sizing Other Than Half-Full

For three-separator other than 50% full of liquid, the following equations are:

Gas Capacity Constraint

Field Units

$$d\,L_{eff} = 420\left(\frac{1-\beta}{1-\alpha}\right)\left(\frac{TZQ_g}{P}\right)\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \tag{18.268}$$

where

$\frac{1-\beta}{1-\alpha} =$ design constant from Figure 18.83a.

SI Units

$$d\,L_{eff} = 34.5\left(\frac{1-\beta}{1-\alpha}\right)\left(\frac{TZQ_g}{P}\right)\left[\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}\right]^{0.5} \tag{18.269}$$

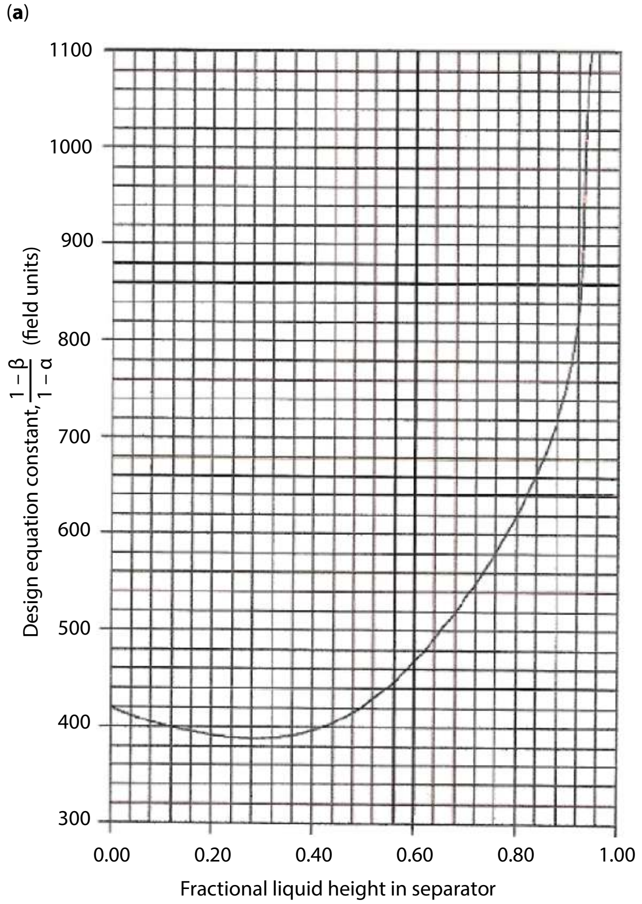


Figure 18.83a Gas capacity constraint design constant [(1 – β)/(1 – α)] vs. liquid height of a cylinder for a horizontal separator other than 50% full of liquid [79].

where

$\frac{1-\beta}{1-\alpha} =$ design constant from Figure 18.83b

Retention Time Constraint

Field Units

$$d^2 L_{eff} = \frac{(t_r)_o Q_o + (t_r)_w Q_w}{1.4\alpha} \tag{18.270}$$

where α = design constant found in Figure 18.84a

SI Units

$$d^2 L_{eff} = 21.0 \frac{(t_r)_o Q_o + (t_r)_w Q_w}{\alpha} \tag{18.271}$$

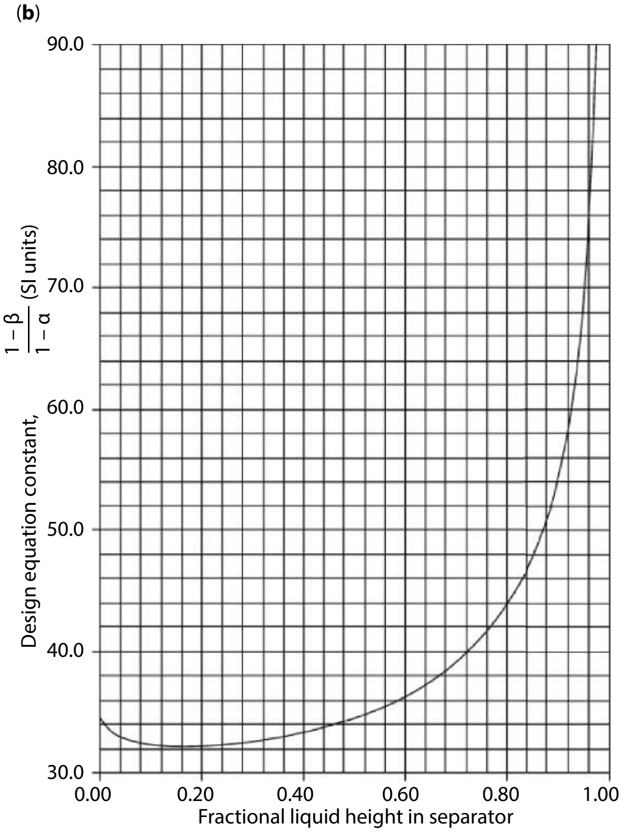


Figure 18.83b Gas capacity constraint design constant [(1 – β)/(1 – α)] vs. liquid height of a cylinder for a horizontal separator other than 50% full of liquid (SI units) [79].

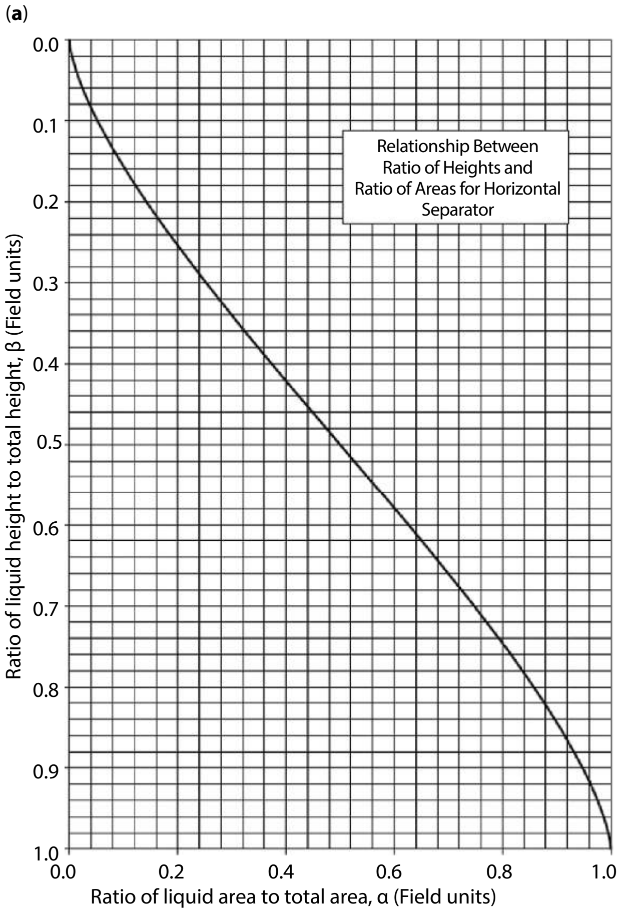


Figure 18.84a Gas capacity constraint design constant [(1 – β)/(1 – α)] vs. liquid height of a cylinder for a horizontal separator other than 50% full of liquid (field units) [79].

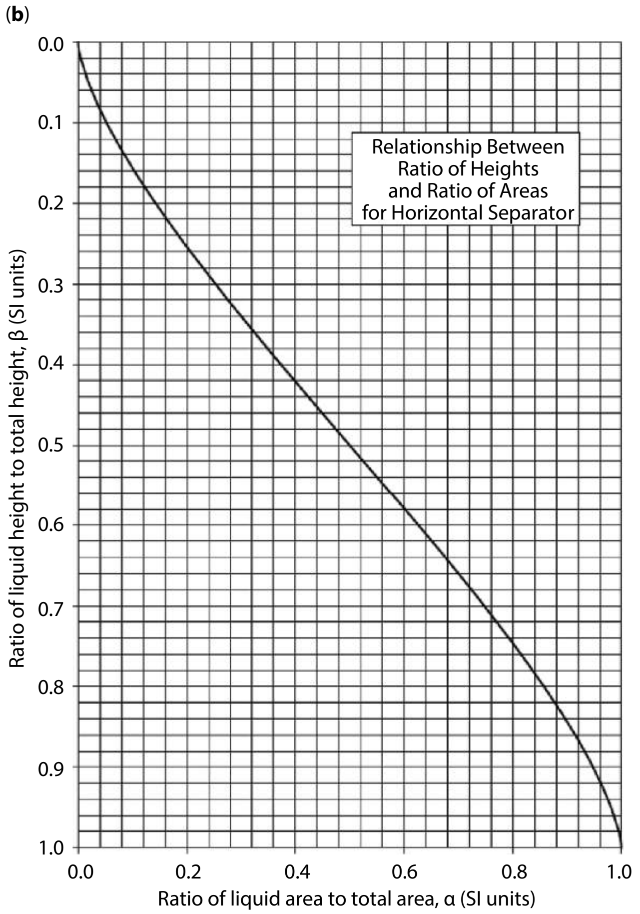


Figure 18.84b Gas capacity constraint design constant [(1 – β)/(1 – α)] vs. liquid height of a cylinder for a horizontal separator other than 50% full of liquid (SI units) [79].

where α = design constant found in Figure 18.84b.

Settling Equation Constraint

From the maximum oil pad thickness, liquid flow rates, and retention times, a maximum vessel diameter may be calculated. The fractional cross-sectional area of the vessel required for water retention may be determined as follows:

$$\alpha_w = \frac{\alpha_l\, Q_w (t_r)_w}{Q_o (t_r)_o + Q_w (t_r)_w} \tag{18.272}$$

where

α_1 = fractional area of liquids
α_w = fractional area of water

The fractional height of the vessel required for the water can be determined by solving the following equation by trial and error:

$$\alpha_w = \frac{1}{80}\cos^{-1}[1-2\beta_w] - \left(\frac{1}{\pi}\right)[1-2\beta_w] \tag{18.273}$$

where β_w represents the fractional height of water.

A maximum vessel diameter may be determined from the fractional heights of the total liquids and water as follows:

$$d_{max} = \frac{(h_o)_{max}}{\beta_l - \beta_w} \qquad (18.274)$$

where d_{max} is the maximum vessel internal diameter in inches (mm). Any vessel diameter less than this maximum may be used to separate specified water droplet size in the specified oil retention time.

Example 18.16 Size a Horizontal Three-Phase Separator (Field Units)

Given:

Q_o = 5000 bpd
Q_w = 3000 bpd
Q_g = 5 MMscfd
P = 100 psia
T = 90°F (550°R)
SG_{oil} = 30°API
SG_w = 1.07
S_g = 0.6
$(t_r)_o = (t_r)_w$ = 10 min
μ_o = 10 cP
μ_w = 1 cP
ρ_g = 0.2975 lb/ft^3
ρ_l = 54.67 lb/ft^3
C_D = 2.211
Z = 0.99

Droplet removal = 100-μm liquids, 500-μm water, 200-μm oil.

Solution

1. Calculate the difference in specific gravities

$$^{o}API = \frac{141.5}{(SG)_o} - 131.5$$

$$(SG)_o = \frac{141.5}{[^{o}API + 131.5]} = \frac{141.5}{[30 + 131.5]} = 0.876$$

$$\Delta SG = 1.07 - 0.876 = 0.194$$

2. Calculate the maximum oil pad thickness $(h_o)_{max}$. Use 500-μm droplet size if no other information is available. From Eq. 18.237

$$(h_o)_{max} = (1.28 \times 10^{-3}) \frac{(10)(0.194)(500^2)}{10}$$
$$= 62.1$$

3. Calculate $\frac{A_w}{A}$ from Eq. 18.252

$$\frac{A_w}{A} = 0.5\frac{(19.8)(10)}{[(33)(10)+(19.8)(10)]}$$
$$= 0.1875.$$

4. Determine β from Figure 18.81. With $A_w/A = 0.1875$, read $\beta = 0.257$
5. Calculate d_{max} from Eq. 18.253.

$$d_{max} = \frac{(h_o)_{max}}{\beta} = \frac{62.1}{0.257} = 241.6\,\text{in.}$$

6. Calculate combinations of d, L_{eff} for $d < d_{max}$ that satisfy the gas capacity constraint. Use 100-μm droplet size if no other information is available. From Eq. 18.208

$$dL_{eff} = 420\left[\frac{(550)(0.99)(5)}{(100)}\right]\left[\left(\frac{0.297}{54.67-0.297}\right)\frac{2.211}{100}\right]^{0.5}$$
$$= 126$$

7. Calculate combinations of d, L_{eff} for $d < d_{max}$ that satisfy the oil and water retention time constraint. From Eq. 18.211 (Table 18.25)

$$d^2L_{eff} = 1.42(10)[3{,}000 + 5{,}000]$$
$$= 113{,}600$$

8. Estimate seam-to-seam length (Table 18.25) from Eq. 18.265 or Eq. 18.266 (Table 18.25).

$$L_{ss} = \frac{4}{3}L_{eff}\ (\text{for liquid capacity})$$
$$L_{ss} = L_{eff} + \frac{d}{12}\ (\text{for gas capacity})$$

9. Select slenderness ratio, SR ($12\,L_{ss}/d$). Choices in the range 3–5 are common (Table 18.25).
10. Choose a reasonable size that does not violate gas capacity restraint oil pad thickness restraint. Possible choices are 72 in., diameter by 30 ft seam-to-seam and 84 in. diameter by 25 ft seam-to-seam length (L_{ss}).

The Excel spreadsheet Example 18.16.xlsx shows the calculations of Example 18.16, and the table below shows the results of the spreadsheet calculations.

Sizing Calculation of a horizontal three-phase separator (Field Units)

Oil flow rate, bpd	5000	bpd
Water flow rate, bpd	3000	bpd
Gas flow rate, MMscfd	5	MMscfd
SG of oil	30	°API
SG of water	1.07	
SG of gas	0.6	
Operating temperature	90	°F
Operating pressure	100	psia
Droplet size removal	140	micron
Retention time, $(t_r)_o = (t_r)_w$	10	min
Compressibility factor, Z	0.99	
Viscosity of oil, μ_o	10	cP
Viscosity of water, μ_w	1	cP
Assume Drag coefficient, C_D	2.01	
Droplet size removal in liquid	100	micron
Droplet size removal in water	500	micron
Droplet size removal in oil	200	micron
Viscosity, μ	0.013	cP
The terminal velocity, V_t	1.3415	ft/s
Reynolds number, Re:	20	
Drag coefficient, C_D:	2.2108	
Maximum pad thickness	62.1	in.
Maximum diameter	241.6	in.
Combination of d and Leff, i.e. dL_{eff}	125.9	

Table 18.25 Horizontal three-phase separator capacity diameter vs. length for retention time constraint $(t_r)_o = (t_r)_w = 10$ min.

d_o (in.)	L_{eff} (ft)	L_{ss} (ft)	$S_R = \left(\frac{12 L_{ss}}{d}\right)$
60	31.6	42.1	8.4
72	21.9	29.2	4.9
84	16.1	21.5	3.1
96	12.3	16.4	2.1
108	9.7	13.0	1.4

Example 18.17 Size a Horizontal Three-Phase Separator (SI Units)

Given:

Q_o = 33 m^3/h
Q_w = 19.8 m^3/h
Q_g = 5902 sm^3/h
P = 6900 kPa
T = 32.2°C (305.4K)
SG_{oil} = 30°API
SG_w = 1.07
S_g = 0.6
$(t_r)_o = (t_r)_w$ = 10 min
μ_o = 10 cP
μ_w = 1 cP
ρ_g = 49 kg/m^3
ρ_l = 866 kg/m^3
C_D = 2.01
Z = 0.99

Droplet removal = 100-μm liquids, 500-μm water, 200-μm oil.

Solution

1. Calculate the difference in specific gravities

$$^{o}\text{API} = \frac{141.5}{(\text{SG})_o} - 131.5$$

$$(\text{SG})_o = \frac{141.5}{[^{o}\text{API} + 131.5]} = \frac{141.5}{[30 + 131.5]} = 0.876$$

$$\Delta\text{SG} = 1.07 - 0.876 = 0.194$$

2. Calculate the maximum oil pad thickness $(h_o)_{max}$. Use 500-μm droplet size if no other information is available. From Eq. 18.249

$$(h_o)_{max} = (0.033)\frac{(10)(0.194)(500^2)}{10}$$
$$= 1600.5$$

3. Calculate $\frac{A_w}{A}$ from Eq. 18.252

$$\frac{A_w}{A} = 0.5\frac{(19.8)(10)}{[(33)(10) + (19.8)(10)]}$$
$$= 0.1875.$$

4. Determine β from Figure 18.81. With A_w/A = 0.1875, read β = 0.257.

5. Calculate d_{max} from Eq. 18.253.

$$d_{max} = \frac{(h_o)_{max}}{\beta} = \frac{1600.5}{0.257} = 6228\,\text{mm}.$$

6. Calculate combinations of d, L_{eff} for $d < d_{max}$ that satisfy the gas capacity constraint. Use 100-μm droplet size if no other information is available. From Eq. 18.209

$$dL_{eff} = 34.5\left[\frac{(305.35)(0.99)(5902)}{(690)}\right]\left[\left(\frac{4.782}{876.16 - 4.782}\right)\frac{2.211}{100}\right]^{0.5}$$
$$= 982.65$$

7. Calculate combinations of d, L_{eff} for $d < d_{max}$ that satisfy the oil and water retention time constraint. From Eq. 18.211 (Table 18.26)

$$d^2L_{eff} = 42{,}000(10)[19.8 + 33]$$
$$= 22{,}176{,}000$$

8. Estimate seam-to-seam length from Eq. 18.265 or Eq. 18.267 (Table 18.26).

$$L_{ss} = \frac{4}{3}L_{eff}\ (\text{for liquid capacity})$$

$$L_{ss} = L_{eff} + \frac{d}{1000}\ (\text{for gas capacity})$$

9. Select slenderness ratio, SR ($1000L_{ss}/d$). Choices in the range 3–5 are common (Table 18.26).
10. Choose a reasonable size that does not violate gas capacity restraint oil pad thickness restraint. Possible choices are 1828 mm, diameter by 8.84 m seam-to-seam and 2133.6 mm diameter by 6.49 m seam-to-seam length (L_{ss}).

Table 18.26 Horizontal three-phase separator capacity diameter vs. length for retention time constraint $(t_r)_o = (t_r)_w = 10$ min.

d (mm)	**L_{eff} (m)**	**L_{ss} (m)**	**$S_R = \left(\frac{1000\,L_{ss}}{d}\right)$**
1524	9.55	12.73	8.4
1828.8	6.63	8.84	4.8
2133.6	4.87	6.49	3.0
2438.4	3.72	4.96	2.0
2743.2	2.95	3.93	1.4

The Excel spreadsheet Example 18.17.xlsx shows the calculations of Example 18.17, and the table below shows the results of the spreadsheet calculations.

Sizing Calculation of a horizontal three-phase separator (S.I. Units)		
Oil flow rate	33	m^3/h
Water flow rate	19.8	m^3/h
Gas flow rate	5902	sm^3/h
SG of oil	30	°API
SG of water	1.07	
SG of gas	0.6	
Operating temperature	32.2	°C
Operating pressure	690	kPa
Droplet size removal	140	micron
Retention time, $(t_r)_o = (t_r)_w$	10	min
Compressibility factor, Z	0.99	
Viscosity of oil, μ_o	10	cP
Viscosity of water, μ_w	1	cP
Assume Drag coefficient, C_D	2.01	
Droplet size removal in liquid	100	micron
Droplet size removal in water	500	micron
Droplet size removal in oil	200	micron
Viscosity, μ	0.013	cP
The terminal velocity, V_t	0.3867	m/s
Reynolds number, Re:	20	
Drag coefficient, C_D:	2.2108	
Maximum oil pad thickness	1600.5	mm
Maximum diameter	6227.6	mm

Example 18.18

A three-phase separation is to be carried out for the well fluid which consists of 110 MMSCFD of gas with 20,000 barrels per day (bpd) of oil and 5000 bpd of water at given specified properties and conditions. The calculation principle is that a further particle of oil should join the oil phase within a specified time due to its vertical upward velocity. Similarly, for the water, the farthest water particle should join the water layer. Determine a preliminary sizing for the horizontal three-phase separator using the data below.

QG	Gas flow rate	MMSCFD	110
QW	Water flow rate	bpd	5000
QO	Oil flow rate	bpd	20,000
G	Specific gravity of gas	From UniSim	0.799
°API	Oil specific gravity	°API (calculated from the density of oil)	42.2
SGW	Water specific gravity	-	1.03
P	Operating pressure	psia	1074.57
T	Operating temperature	°R	570
DM	Diameter of water droplet to be separated from oil	microns (as per specification)	500
DML	Diameter of liquid (water and oil) droplet to be separated from gas	microns (as per specification)	150
TRW	Water retention time	min (as per selection)	5/10
TRO	Oil retention time	min (as per selection)	5/10
LVIS	Viscosity of liquid	cP	1.433

Solution

An Excel spreadsheet (Example 18.18.xlsx) has been developed for Example 18.8. Details of the calculations are shown in the spreadsheet and Table 18.27a. Table 18.27b shows the results of liquid capacity constraint at retention time of 5 and 10 min, respectively. From the calculations, 102 in. diameter and a seam-to-seam length of 25 ft are suitable for the separator. Concerning oil water separation, if an emulsion occurs, and consequently 5 min residence time is inadequate, then either the retention time can be increased or coalescer pad consisting of parallel plates can be designed and fitted into the separator so that the same effectiveness is attained. This may work for particles up to ~50% size of what was assumed for separation. For smaller particles, a reliable method will be to fit proprietary coalescers as recommended by vendors based on particle size distribution.

The Excel spreadsheet calculation allows the designer to vary the diameter and trade-offs between length and diameter from which a desired vessel is selected. Increasing the diameter can result in a step change of the shell thickness of the vessel, which increases the cost of the vessel. In the case of offshore platforms, it could result in the cost of

Table 18.27a Horizontal three-phase capacity diameter vs. length for liquid retention time $(t_r)_o = (t_r)_w = 5$ min.

Vessel diameter d, in.	D, ft	L_{eff} (vessel effect length), ft	L_{ss} (seam-to-seam length), ft	SR, slenderness ratio ($12*L_{ss}/d$)
50	4.2	71	94.67	22.72
62	5.2	46.18	61.57	11.92
74	6.2	32.41	43.21	7.01
86	7.2	24	32	4.47
98	8.2	18.48	24.64	3.02
102	8.5	17.06	22.75	2.68
124	10.3	11.54	15.39	1.49

Table 18.27b Horizontal three-phase capacity diameter vs. length for liquid retention time $(t_r)_o = (t_r)_w = 10$ min.

Vessel diameter d, in.	D, ft	L_{eff} (vessel effect length), ft	L_{ss} (seam-to-seam length), ft	SR, slenderness ratio ($12*L_{ss}/d$)
50	4.2	142	189.33	45.44
62	5.2	92.35	123.13	23.83
74	6.2	64.83	86.44	14.02
86	7.2	48	64	8.93
98	8.2	36.96	49.28	6.03
102	8.5	34.12	45.49	5.35
124	10.3	23.09	30.79	2.98
136	11.3	19.19	25.59	2.26

the platform structure. The designer in this case should optimize the weight at the same time and it should result in utilizing the space, or result in a utilizable length on the platform. This means that in smaller platforms, it would be advisable to use either the entire length or the entire width of the platform, leaving only enough space for operation and maintenance [85].

While locating a separator on an offshore platform, the general relationship between hazard and layout should be as follows [86]:

- Make the initial layout.
- Visualize hazard events.
- Study the release and dispersion of gas.
- Consider chances of fire/explosion.
- Consider chances of effect on facility.
- Modify the layout to minimize damage.

If the separator is to be located on floating platforms such as Floating Production Storage and Offloading (FPSO) unit, the sizes have to be larger to take care of the natural rolling and pitching movements of the FPSO unit. It may be noted that FPSO units are production platforms located at sea but not firmly supported from the sea bed.

The Excel spreadsheet Example 18.18.xlsx and worksheets (Example 18-18-1 and 18-18-2, with $(t_r)_o = (t_r)_w = 5$ and 10 mins. respectively) show the calculations of Example 18.18.

A Case Study (UniSim Design)

Crude oil produced from oil reservoirs contains water, gas, salt, sediment and small amounts of metal is to be separated in a three-phase horizontal separator. The components of the crude and stream parameters are as follows:

Stream	Oil sea line
Vapor/Phrase Fraction	0.0848
Temperature, °C	26.32
Pressure, kPa	1200
Molar flow (kg mole/h)	2511
Mass flow (kg/h)	197,489.7317

Nitrogen (mole fraction)	0.002992
H_2S	0
CO_2	0.006781
H_2O	0.300957
Methane	0.087256
Ethane	0.037495
Propane	0.039689
i-Butane	0.012565
n-Butane	0.026426
i-Pentane	0.016254
n-Pentane	0.021739
n-Hexane	0.042780
n-Heptane	0.057738
n-Octane	0.039789
n-Nonane	0.036598
n-Decane	0.034105
n-C_{11}	0.024432
n-C_{12}	0.212405

Solution

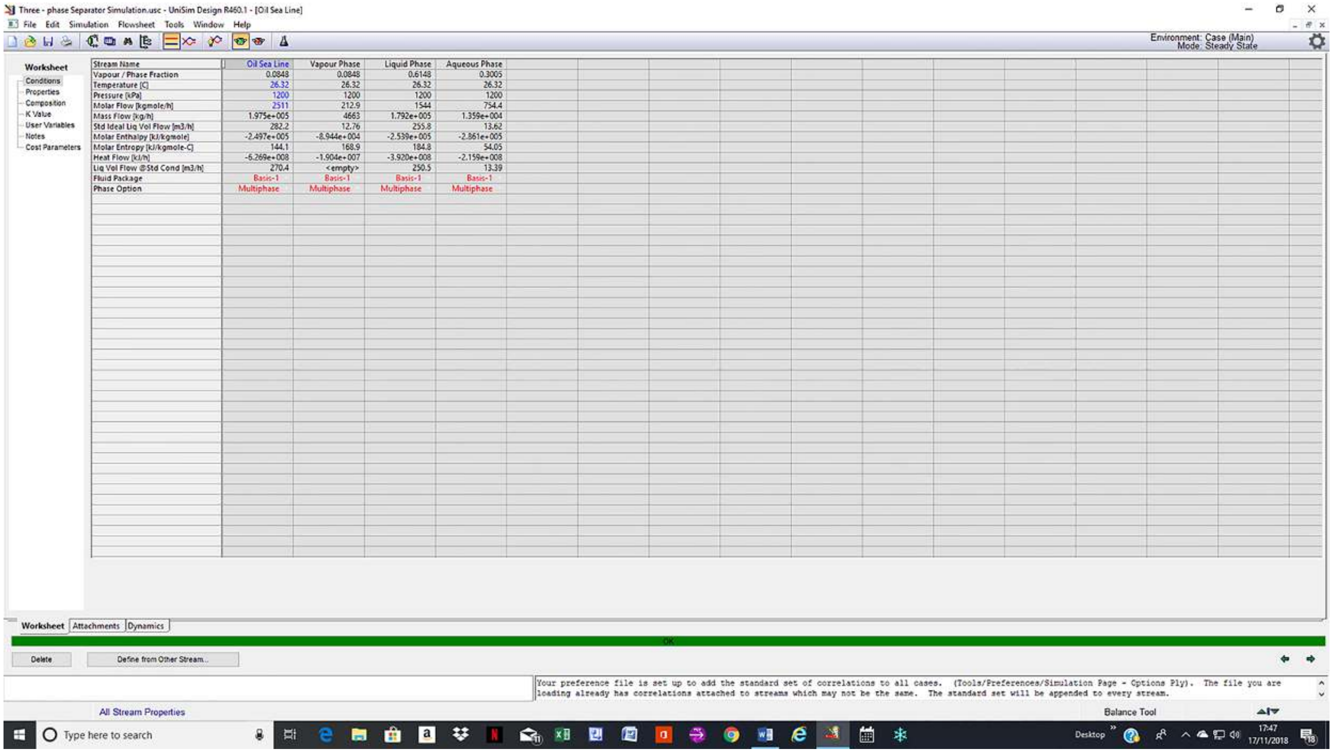
UniSim Design R460.1® simulator is used to carry out separation of crude oil in a horizontal three-phase separator. The following Figures 18.85–18.88 show the input of the data in various corresponding windows and parameter pages and Figure 18.88 shows the process flow diagram of the converged simulation. Table 18.28 shows the results of the simulation can be accessed from Three-phase Separator Simulation.usc.

Rules of Thumb: Tanks and Vessels

In general, vessels are essential for the smooth running of plants as they pertain to storage and processing. Process engineers should strive to minimize the size of vessels, while maintaining the desired plant operation, as this falls into two categories, namely: safety and cost.

Safety of vessels involves the integrity of the tank, its ability to contain its fluids encountering process upsets, degradation of the tank with respect to corrosion and erosion, physical damage and external events due to fire. Another essential consideration is in safely accessing the inside of the vessel for inspection or cleaning and maintenance after it is installed. A greater concern to operators/technicians is asphyxiation as well positioned access flanges should be provided along with a means to ventilate the tank interior and install personnel harnesses. An aspect of tank safety relates to an inherent safer tank where inventory in the tanks is limited/controlled either for storage or for further processing of fluids.

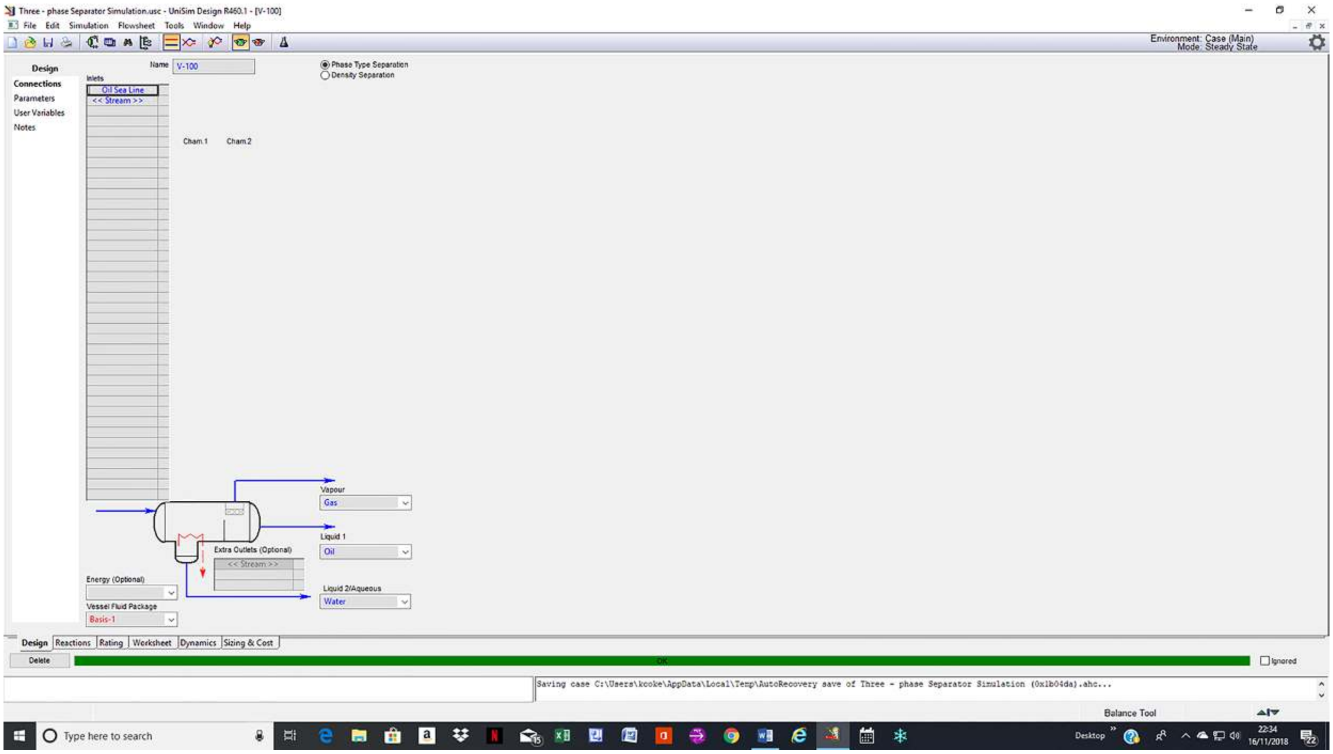
For a larger tank, fabrication and installation are preferably carried out at the plant site.



Stream Name	Oil Sea Line	Vapour Phase	Liquid Phase	Aqueous Phase
Vapour / Phase Fraction	0.0848	0.0848	0.6148	0.3005
Temperature [C]	26.32	26.32	26.32	26.32
Pressure [kPa]	1200	1200	1200	1200
Molar Flow [kgmole/h]	2511	212.9	1544	754.4
Mass Flow [kg/h]	1.975e+005	4663	1.792e+005	1.359e+004
Std Ideal Liq Vol Flow [m3/h]	282.2	12.76	255.8	13.62
Molar Enthalpy [kJ/kgmole]	-2.497e+005	-8.944e+004	-2.539e+005	-2.861e+005
Molar Entropy [kJ/kgmole-C]	144.1	168.9	184.8	54.05
Heat Flow [kJ/h]	-6.269e+008	-1.904e+007	-3.920e+008	-2.159e+008
Liq Vol Flow @Std Cond [m3/h]	270.4	<empty>	250.5	13.39
Fluid Package	Basis-1	Basis-1	Basis-1	Basis-1
Phase Option	Multiphase	Multiphase	Multiphase	Multiphase

```
Your preference file is set up to add the standard set of correlations to all cases.  (Tools/Preferences/Simulation Page - Options Ply).  The file you are
loading already has correlations attached to streams which may not be the same.  The standard set will be appended to every stream.
```

Figure 18.85 A snapshot of the Worksheet window showing the Conditions page (courtesy of Honeywell Process Solution. UniSim Design R460.1, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



Name: V-100

Inlets: Oil Sea Line, << Stream >>

● Phase Type Separation
○ Density Separation

Vapour: Gas

Liquid 1: Oil

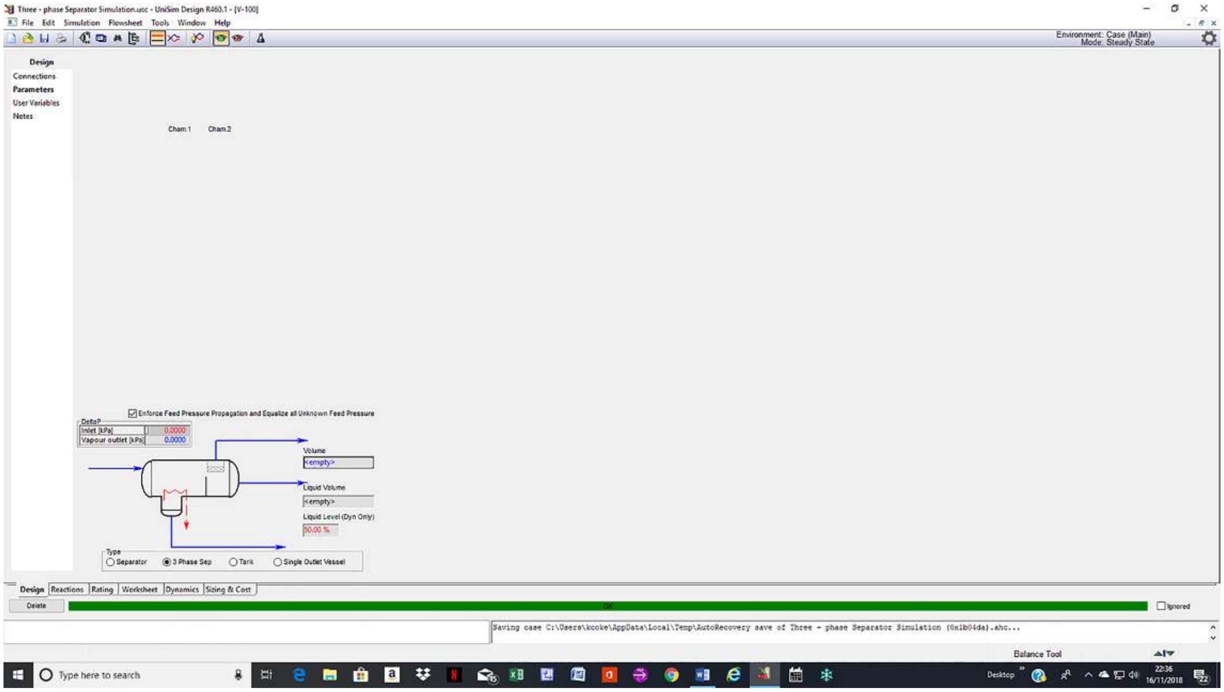
Liquid 2/Aqueous: Water

Energy (Optional):

Vessel Fluid Package: Basis-1

```
Saving case C:\Users\kooke\AppData\Local\Temp\AutoRecovery save of Three - phase Separator Simulation (0x1b04da).ahc...
```

Figure 18.86 A snapshot of the Design window showing the Connections page (courtesy of Honeywell Process Solution. UniSim Design R460.1, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



☑ Enforce Feed Pressure Propagation and Equalize all Unknown Feed Pressure

DeltaP
Inlet [kPa] 0.0000
Vapour outlet [kPa] 0.0000

Volume <empty>
Liquid Volume <empty>
Liquid Level (Dyn Only) 50.00 %

Type ○ Separator ● 3 Phase Sep ○ Tank ○ Single Outlet Vessel

```
Saving case C:\Users\kcoke\AppData\Local\Temp\AutoRecovery save of Three - phase Separator Simulation (0x1b04da).ahc...
```

Figure 18.87 A snapshot of the Design window showing the Parameters page (courtesy of Honeywell Process Solution. UniSim Design R460.1, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).

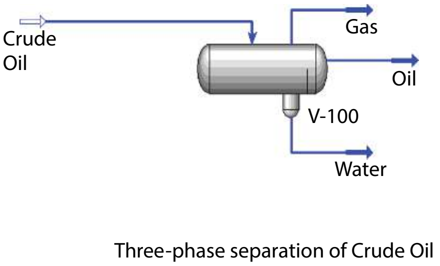
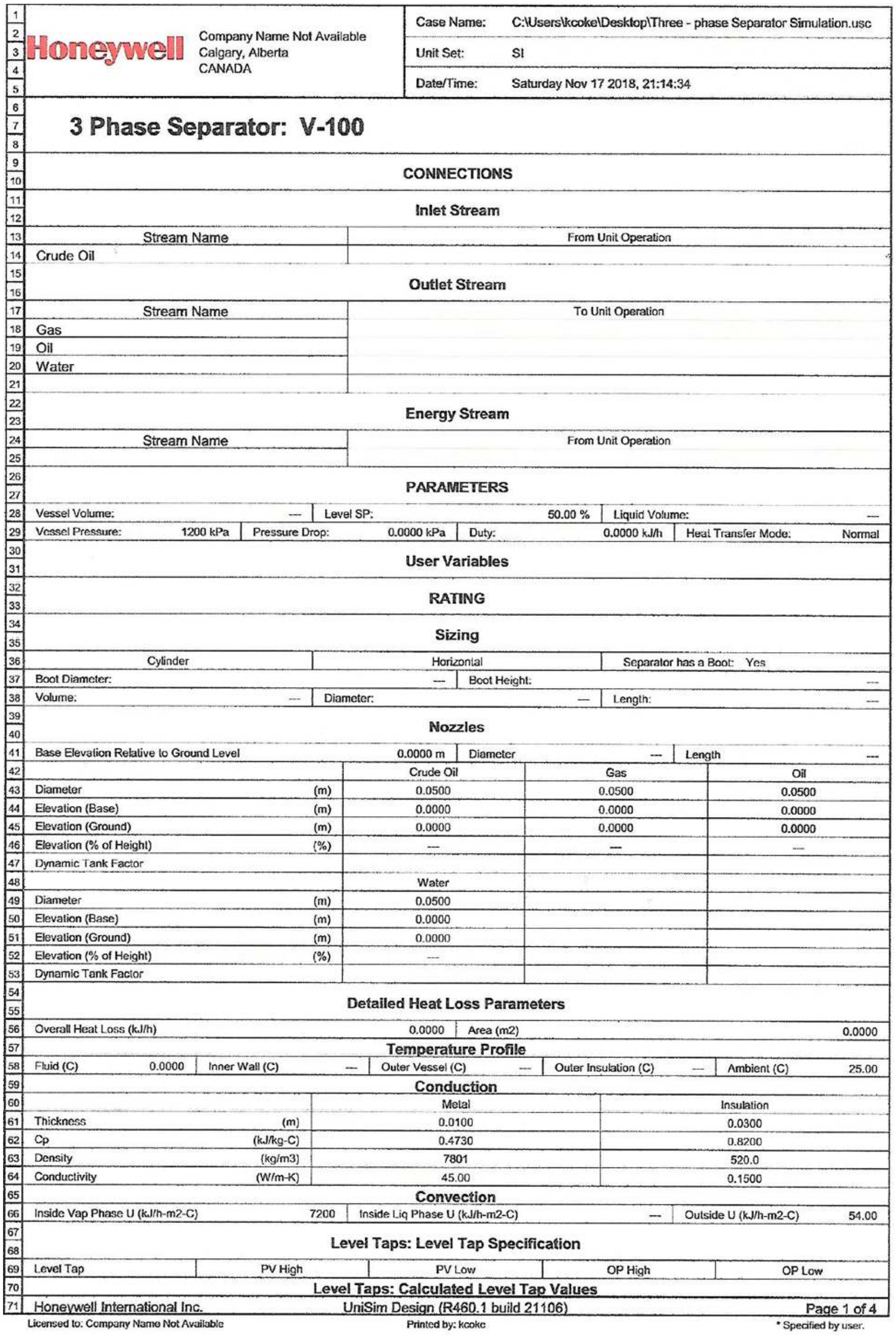


Figure 18.88 A snapshot of the process flow diagram (courtesy of Honeywell Process Solution. UniSim Design R460.1, Honeywel®, and UniSim® are registered trademarks of Honeywell International Inc.).

Inventory held in the tanks for raw materials, intermediates or final product is an operating cost that can significantly affect the cost of the products, as other operating costs are affected by vessel size, including operator hours, cooling water, steam and power. Large atmospheric storage tanks as employed at oil refineries facilities are designed by specialists who follow strict guidance and codes. These sizes range up to 200 m^3 with diameters reaching 20 m. Smaller size less than 60,000 liters atmospheric tanks and pressure related vessels are often specified by process and mechanical engineers. The following rules of thumb may be useful [87]:

- They must be designed, fabricated and tested in accordance with applicable codes.
- The user is responsible for specifying loadings that are used to calculate the vessel wall, thickness and reinforcements. Factors include: internal/external pressure, ambient and operational temperatures; static pressure and mass of contents in operating and test conditions; wind and earthquake conditions; reaction forces and moments resulting from supports, attachments, piping, thermal expansion, etc., corrosion; fatigue, and decomposition of unstable fluids.

Table 18.28 Results of the three-phase separator simulation.



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	C:\Users\kcoke\Desktop\Three - phase Separator Simulation.usc
Unit Set:	SI
Date/Time:	Saturday Nov 17 2018, 21:14:34

3 Phase Separator: V-100

CONNECTIONS

Inlet Stream

Stream Name	From Unit Operation
Crude Oil	

Outlet Stream

Stream Name	To Unit Operation
Gas	
Oil	
Water	

Energy Stream

Stream Name	From Unit Operation

PARAMETERS

Vessel Volume:	---	Level SP:	50.00 %	Liquid Volume:	---
Vessel Pressure: 1200 kPa	Pressure Drop: 0.0000 kPa	Duty:	0.0000 kJ/h	Heat Transfer Mode:	Normal

User Variables

RATING

Sizing

Cylinder	Horizontal	Separator has a Boot: Yes
Boot Diameter: ---	Boot Height: ---	
Volume: ---	Diameter: ---	Length: ---

Nozzles

Base Elevation Relative to Ground Level 0.0000 m	Diameter ---	Length ---

		Crude Oil	Gas	Oil
Diameter	(m)	0.0500	0.0500	0.0500
Elevation (Base)	(m)	0.0000	0.0000	0.0000
Elevation (Ground)	(m)	0.0000	0.0000	0.0000
Elevation (% of Height)	(%)	---	---	---
Dynamic Tank Factor				
		Water		
Diameter	(m)	0.0500		
Elevation (Base)	(m)	0.0000		
Elevation (Ground)	(m)	0.0000		
Elevation (% of Height)	(%)	---		
Dynamic Tank Factor				

Detailed Heat Loss Parameters

Overall Heat Loss (kJ/h)	0.0000	Area (m2)	0.0000

Temperature Profile

Fluid (C)	0.0000	Inner Wall (C)	---	Outer Vessel (C)	---	Outer Insulation (C)	---	Ambient (C)	25.00

Conduction

		Metal	Insulation
Thickness	(m)	0.0100	0.0300
Cp	(kJ/kg-C)	0.4730	0.8200
Density	(kg/m3)	7801	520.0
Conductivity	(W/m-K)	45.00	0.1500

Convection

Inside Vap Phase U (kJ/h-m2-C)	7200	Inside Liq Phase U (kJ/h-m2-C)	---	Outside U (kJ/h-m2-C)	54.00

Level Taps: Level Tap Specification

Level Tap	PV High	PV Low	OP High	OP Low

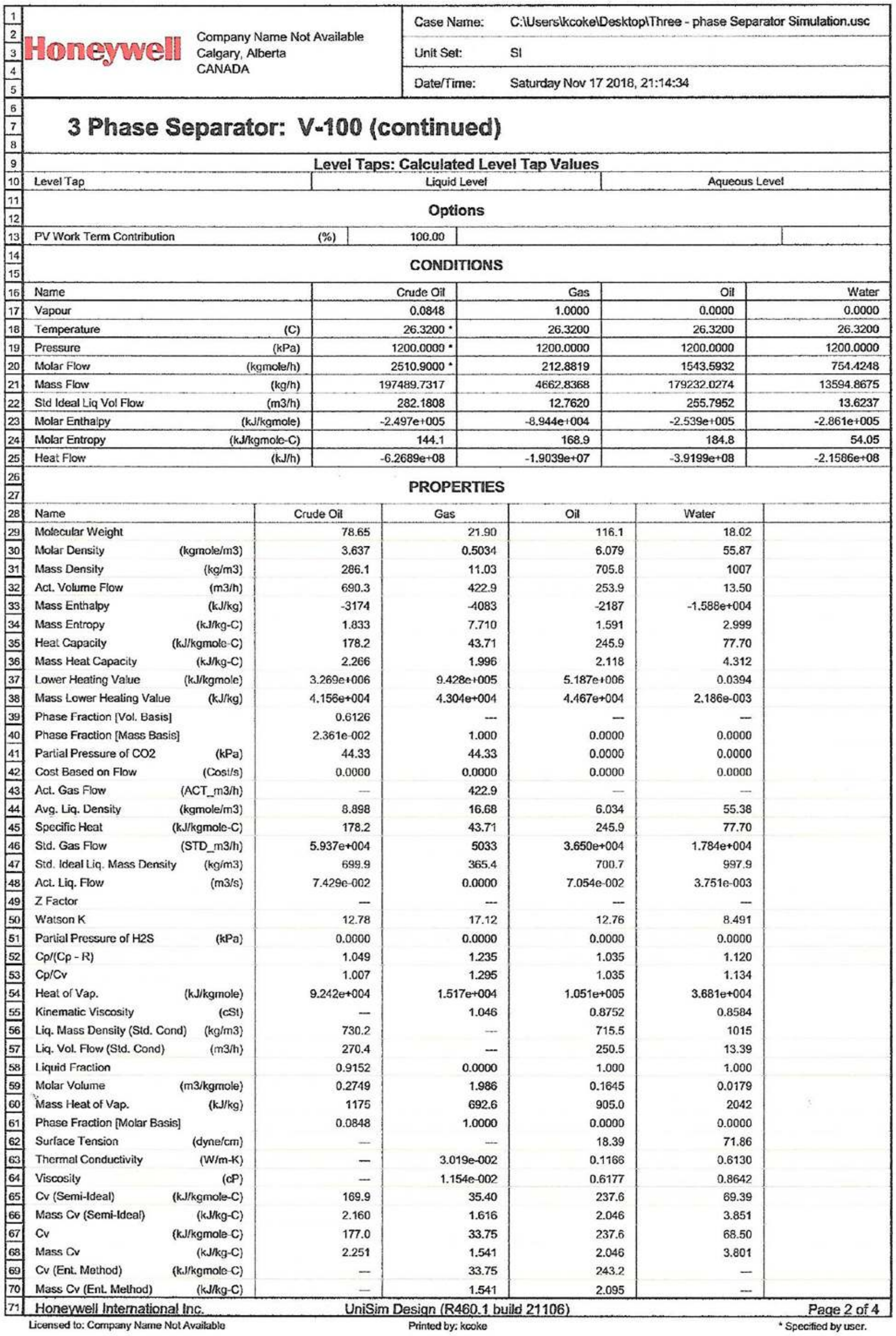
Level Taps: Calculated Level Tap Values

Honeywell International Inc. UniSim Design (R460.1 build 21106) Page 1 of 4

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

(*Continued*)

Table 18.28 Results of the three-phase separator simulation. (*Contnued*)



Honeywell
Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	C:\Users\kcoke\Desktop\Three - phase Separator Simulation.usc
Unit Set:	SI
Date/Time:	Saturday Nov 17 2018, 21:14:34

3 Phase Separator: V-100 (continued)

Level Taps: Calculated Level Tap Values

Level Tap	Liquid Level	Aqueous Level

Options

PV Work Term Contribution	(%)	100.00		

CONDITIONS

Name		Crude Oil	Gas	Oil	Water
Vapour		0.0848	1.0000	0.0000	0.0000
Temperature	(C)	26.3200 *	26.3200	26.3200	26.3200
Pressure	(kPa)	1200.0000 *	1200.0000	1200.0000	1200.0000
Molar Flow	(kgmole/h)	2510.9000 *	212.8819	1543.5932	754.4248
Mass Flow	(kg/h)	197489.7317	4662.8368	179232.0274	13594.8675
Std Ideal Liq Vol Flow	(m3/h)	282.1808	12.7620	255.7952	13.6237
Molar Enthalpy	(kJ/kgmole)	-2.497e+005	-8.944e+004	-2.539e+005	-2.861e+005
Molar Entropy	(kJ/kgmole-C)	144.1	168.9	184.8	54.05
Heat Flow	(kJ/h)	-6.2689e+08	-1.9039e+07	-3.9199e+08	-2.1586e+08

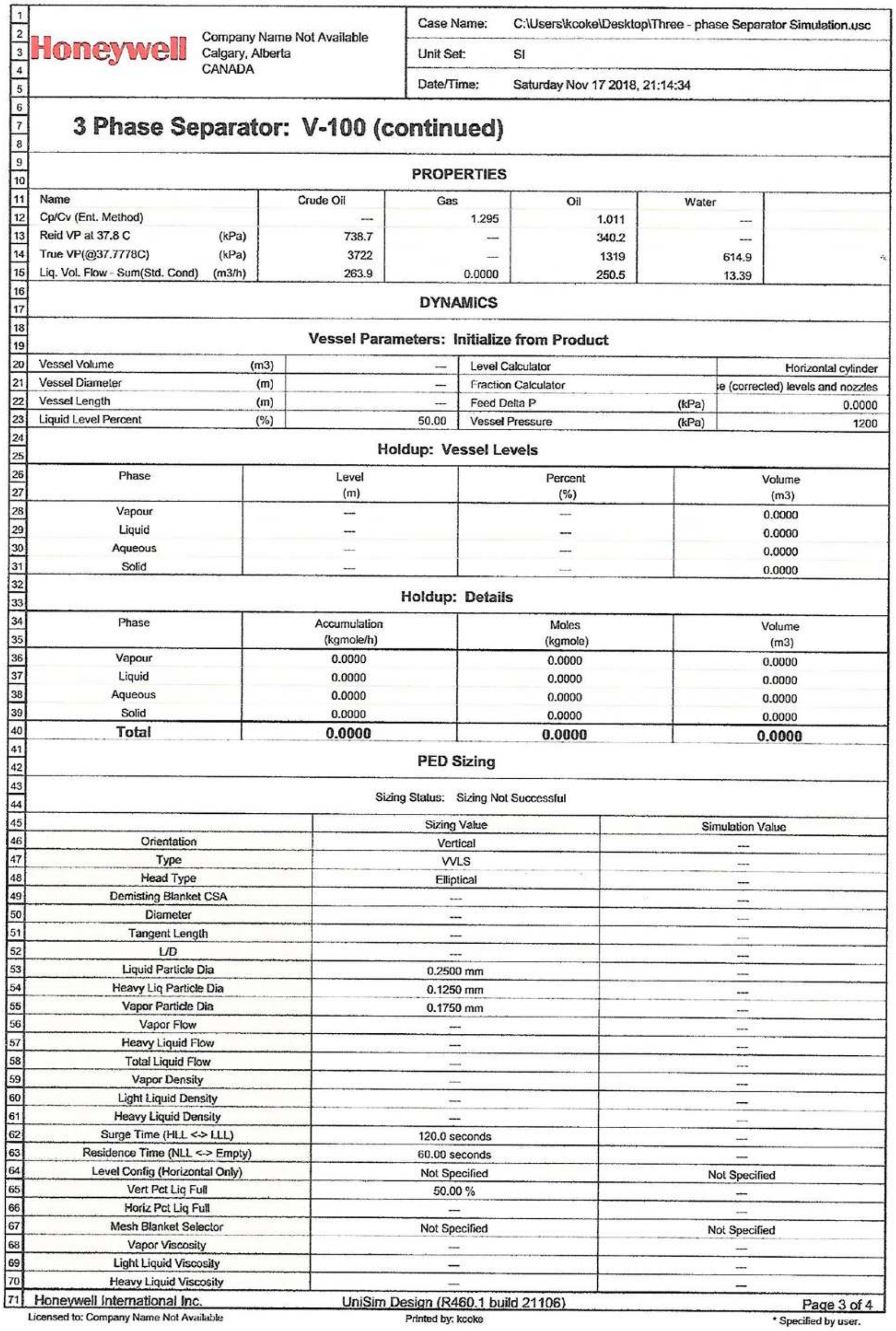
PROPERTIES

Name		Crude Oil	Gas	Oil	Water	
Molecular Weight		78.65	21.90	116.1	18.02	
Molar Density	(kgmole/m3)	3.637	0.5034	6.079	55.87	
Mass Density	(kg/m3)	286.1	11.03	705.8	1007	
Act. Volume Flow	(m3/h)	690.3	422.9	253.9	13.50	
Mass Enthalpy	(kJ/kg)	-3174	-4083	-2187	-1.588e+004	
Mass Entropy	(kJ/kg-C)	1.833	7.710	1.591	2.999	
Heat Capacity	(kJ/kgmole-C)	178.2	43.71	245.9	77.70	
Mass Heat Capacity	(kJ/kg-C)	2.266	1.996	2.118	4.312	
Lower Heating Value	(kJ/kgmole)	3.269e+006	9.428e+005	5.187e+006	0.0394	
Mass Lower Heating Value	(kJ/kg)	4.156e+004	4.304e+004	4.467e+004	2.186e-003	
Phase Fraction [Vol. Basis]		0.6126	---	---	---	
Phase Fraction [Mass Basis]		2.361e-002	1.000	0.0000	0.0000	
Partial Pressure of CO2	(kPa)	44.33	44.33	0.0000	0.0000	
Cost Based on Flow	(Cost/s)	0.0000	0.0000	0.0000	0.0000	
Act. Gas Flow	(ACT_m3/h)	---	422.9	---	---	
Avg. Liq. Density	(kgmole/m3)	8.898	16.68	6.034	55.38	
Specific Heat	(kJ/kgmole-C)	178.2	43.71	245.9	77.70	
Std. Gas Flow	(STD_m3/h)	5.937e+004	5033	3.650e+004	1.784e+004	
Std. Ideal Liq. Mass Density	(kg/m3)	699.9	365.4	700.7	997.9	
Act. Liq. Flow	(m3/s)	7.429e-002	0.0000	7.054e-002	3.751e-003	
Z Factor		---	---	---	---	
Watson K		12.78	17.12	12.76	8.491	
Partial Pressure of H2S	(kPa)	0.0000	0.0000	0.0000	0.0000	
Cp/(Cp - R)		1.049	1.235	1.035	1.120	
Cp/Cv		1.007	1.295	1.035	1.134	
Heat of Vap.	(kJ/kgmole)	9.242e+004	1.517e+004	1.051e+005	3.681e+004	
Kinematic Viscosity	(cSt)	---	1.046	0.8752	0.8584	
Liq. Mass Density (Std. Cond)	(kg/m3)	730.2	---	715.5	1015	
Liq. Vol. Flow (Std. Cond)	(m3/h)	270.4	---	250.5	13.39	
Liquid Fraction		0.9152	0.0000	1.000	1.000	
Molar Volume	(m3/kgmole)	0.2749	1.986	0.1645	0.0179	
Mass Heat of Vap.	(kJ/kg)	1175	692.6	905.0	2042	
Phase Fraction [Molar Basis]		0.0848	1.0000	0.0000	0.0000	
Surface Tension	(dyne/cm)	---	---	18.39	71.86	
Thermal Conductivity	(W/m-K)	---	3.019e-002	0.1166	0.6130	
Viscosity	(cP)	---	1.154e-002	0.6177	0.8642	
Cv (Semi-Ideal)	(kJ/kgmole-C)	169.9	35.40	237.6	69.39	
Mass Cv (Semi-Ideal)	(kJ/kg-C)	2.160	1.616	2.046	3.851	
Cv	(kJ/kgmole-C)	177.0	33.75	237.6	68.50	
Mass Cv	(kJ/kg-C)	2.251	1.541	2.046	3.801	
Cv (Ent. Method)	(kJ/kgmole-C)	---	33.75	243.2	---	
Mass Cv (Ent. Method)	(kJ/kg-C)	---	1.541	2.095	---	

Honeywell International Inc. UniSim Design (R460.1 build 21106) Page 2 of 4
Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

(*Continued*)

Table 18.28 Results of the three-phase separator simulation. (*Contnued*)



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	C:\Users\kcoke\Desktop\Three - phase Separator Simulation.usc
Unit Set:	SI
Date/Time:	Saturday Nov 17 2018, 21:14:34

3 Phase Separator: V-100 (continued)

PROPERTIES

Name		Crude Oil	Gas	Oil	Water	
Cp/Cv (Ent. Method)		---	1.295	1.011	---	
Reid VP at 37.8 C	(kPa)	738.7	---	340.2	---	
True VP(@37.7778C)	(kPa)	3722	---	1319	614.9	
Liq. Vol. Flow - Sum(Std. Cond)	(m3/h)	263.9	0.0000	250.5	13.39	

DYNAMICS

Vessel Parameters: Initialize from Product

Vessel Volume	(m3)	---	Level Calculator		Horizontal cylinder
Vessel Diameter	(m)	---	Fraction Calculator		se (corrected) levels and nozzles
Vessel Length	(m)	---	Feed Delta P	(kPa)	0.0000
Liquid Level Percent	(%)	50.00	Vessel Pressure	(kPa)	1200

Holdup: Vessel Levels

Phase	Level (m)	Percent (%)	Volume (m3)
Vapour	---	---	0.0000
Liquid	---	---	0.0000
Aqueous	---	---	0.0000
Solid	---	---	0.0000

Holdup: Details

Phase	Accumulation (kgmole/h)	Moles (kgmole)	Volume (m3)
Vapour	0.0000	0.0000	0.0000
Liquid	0.0000	0.0000	0.0000
Aqueous	0.0000	0.0000	0.0000
Solid	0.0000	0.0000	0.0000
Total	**0.0000**	**0.0000**	**0.0000**

PED Sizing

Sizing Status: Sizing Not Successful

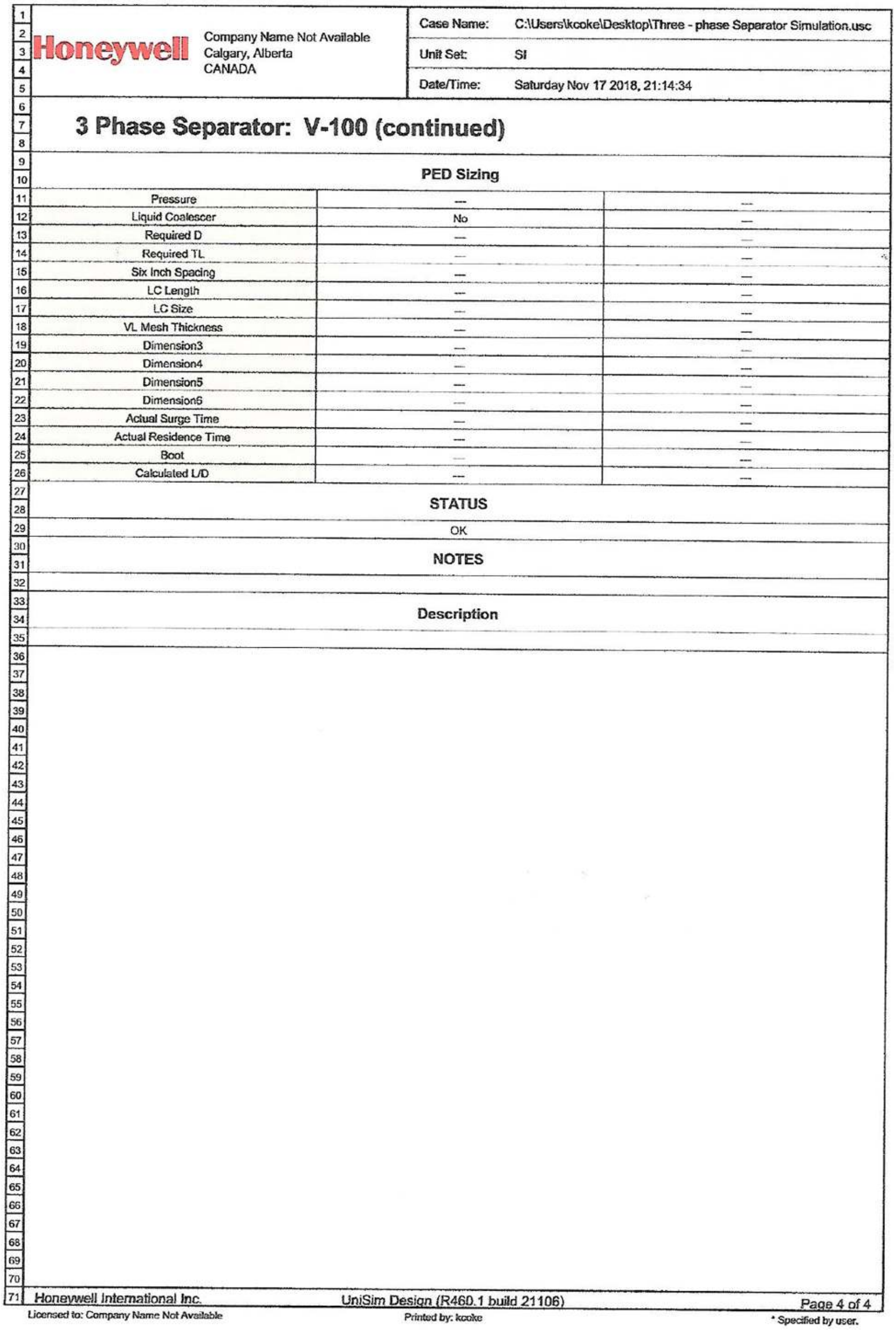
	Sizing Value	Simulation Value
Orientation	Vertical	---
Type	VVLS	---
Head Type	Elliptical	---
Demisting Blanket CSA	---	---
Diameter	---	---
Tangent Length	---	---
L/D	---	---
Liquid Particle Dia	0.2500 mm	---
Heavy Liq Particle Dia	0.1250 mm	---
Vapor Particle Dia	0.1750 mm	---
Vapor Flow	---	---
Heavy Liquid Flow	---	---
Total Liquid Flow	---	---
Vapor Density	---	---
Light Liquid Density	---	---
Heavy Liquid Density	---	---
Surge Time (HLL <-> LLL)	120.0 seconds	---
Residence Time (NLL <-> Empty)	60.00 seconds	---
Level Config (Horizontal Only)	Not Specified	Not Specified
Vert Pct Liq Full	50.00 %	---
Horiz Pct Liq Full	---	---
Mesh Blanket Selector	Not Specified	Not Specified
Vapor Viscosity	---	---
Light Liquid Viscosity	---	---
Heavy Liquid Viscosity	---	---

Honeywell International Inc. UniSim Design (R460.1 build 21106) Page 3 of 4

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

(*Continued*)

Table 18.28 Results of the three-phase separator simulation. (*Contnued*)



Honeywell	Company Name Not Available Calgary, Alberta CANADA
Case Name:	C:\Users\kcoke\Desktop\Three - phase Separator Simulation.usc
Unit Set:	SI
Date/Time:	Saturday Nov 17 2018, 21:14:34

3 Phase Separator: V-100 (continued)

PED Sizing

Pressure	—	—
Liquid Coalescer	No	—
Required D	—	—
Required TL	—	—
Six Inch Spacing	—	—
LC Length	—	—
LC Size	—	—
VL Mesh Thickness	—	—
Dimension3	—	—
Dimension4	—	—
Dimension5	—	—
Dimension6	—	—
Actual Surge Time	—	—
Actual Residence Time	—	—
Boot	—	—
Calculated L/D	—	—

STATUS

OK

NOTES

Description

Honeywell International Inc. UniSim Design (R460.1 build 21106) Page 4 of 4

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

- The aspect ratio (vertical straight-side height divided by diameter) is usually between 1:1 to 1.5:1. Taller vessels with aspect ratio ranging to about 4:1, are used when necessary to maximize heat transfer through a jacket, to maximize contact time of a sparged gas, or for other process requirements.
- ASME F & D heads (torispherical) are usually specified for pressures to 20 bar. Ellipsoidal (2:1) heads are used for pressures from 20-100 bar. Very high pressure applications, above 100 bar, utilize hemispherical heads. Conical bottoms are used for some crystallizer when it is desired that precipiates flow freely to the bottom outlet nozzle.
- The working volume of an agitated vessel should be about 80% of the volume measured at the top tangent line. If the aspect ratio exceeds 2:1, then multiple impellers may be needed.
- Determining heating and cooling duty using factors including control of process exotherms, heat up and cool-down loads (time-based), boiling, thermal losses to the environment, and heat input from agitators and pumps.

Spherical Separators

Spherical separators are vessels where there is no cylindrical shell between the two heads. Fluids enters the vessel through the inlet diverter where the flow stream is split into two streams. Liquid falls to the liquid collection section through openings in a horizontal plate located slightly below the gas–liquid interface. The thin liquid layer across the plate makes it easier for any entrained gases to separate and rise to the gravity settling section. Gases rising out of the liquids pass through the mist extractor and out of the separator through the gas outlet. Liquid level is maintained by a float connected to a dump valve. Pressure is maintained by a back-pressure control valve while the liquid level is maintained by a liquid dump valve. These separator types are very difficult to size and operate. However, they may be very efficient from a pressure containment standpoint, but because they have limited liquid surge capability and also exhibit fabrication difficulties, they are seldom employed in oil field facilities. Figure 18.89 shows a schematic of a spherical separator and Table 18.13 shows the features and differences of vertical, horizontal and spherical separators.

Operating Problems

The operation of separators presents the following problems:

- Foamy crude
- Paraffin

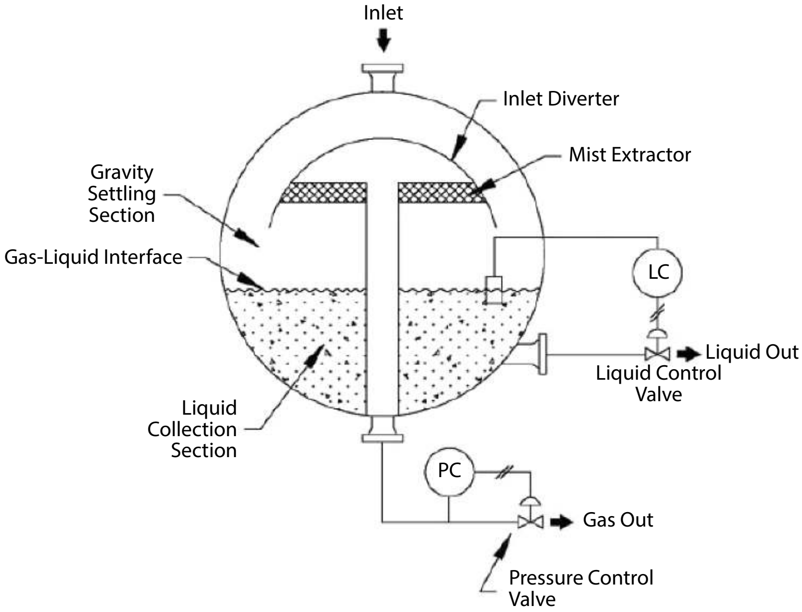


Figure 18.89 Schematic of a spherical separator [79].

- Sand
- Liquid Carryover
- Gas blowby
- Emulsions

Foamy Crude

The major cause of foam in crude oil other than water is the appearance of impurities. These are difficult to remove before the stream reaches the separator. An impurity that causes foam is carbon dioxide (CO_2). Foam presents no problem within a separator if the internal design ensures adequate time or sufficient coalescing surface for the foam to "break." Problems of foaming in a separator are:

- Mechanical control of liquid level is aggravated because any control device must deal with three liquid phases, an emulsion is the third phase and instead of two-phases.
- Foam has a large volume-to-weight ratio. Therefore, it can occupy much of the vessel space that would otherwise be available in the liquid-collecting or gravity settling sections.
- In an uncontrolled foam bank, it becomes impossible to remove separated gas or degassed oil from the vessel without entraining some of the foamy material in either the liquid or the gas outlets.

As the foam is dispersed, it creates very small liquid droplets, which carry over. The amount of foam is dependent on the pressure drop to which the inlet liquid is subjected, as well as the characteristics of the liquid as separator conditions. In some cases, the effect of temperature may be significant. Changing the temperature at which a foamy oil is separated has two effects on the foam. These are the change in the viscosity of oil, as an increase in temperature decreases the oil viscosity thereby making it easier for the rage to escape from the oil. The other is the change in gas-oil equilibrium. A temperature increase will increase the amount of gas that escapes from the oil. Generally, it is difficult to predict the effects of temperature on the foaming tendencies of an oil. Observations are that for low API gravity crude (heavy oils), with low gas-to-oil ratios (GORs), increasing the temperature decreases the foaming tendency of oils. Correspondingly, for high API crude (light oils), with high GORs, increasing the operating temperature decreases the oils' foaming tendencies. However, increasing the operating temperature for a high API gravity crude (light oil) with low GORs may increase the foaming tendencies. Oils in this category are rich in intermediates that have a tendency to evolve to the gas phase as the temperature increases [79].

Foam will often be effective by increasing the capacity of a given separator. Foam can be reduced by:

- *Using a defoaming pack.*
- *Using depressant chemicals*
- *Utilizing heat to break it down.*

Paraffin

Coalescing plates in the liquid section and mesh pad mist extractors in the gas section are particularly prone to plugging by accumulation of paraffin waxes. Using centrifugal mist extractors or plate may be considered where it is determined that paraffin presents a potential problem. Manways, handholes, and nozzles should be provided to allow steam, solvent, or other types of cleaning of the separator internals. Furthermore, the bulk temperature of the liquid should always be kept above the cloud point of the crude oil to prevent paraffin wax formation in the separators.

Sand

Sand in separators causes cutout of valve trim, plugging of separator internals and accumulation in the bottom of the vessel that results in level control problems. Special hard trim can minimize the effects of sand on the valves as accumulation

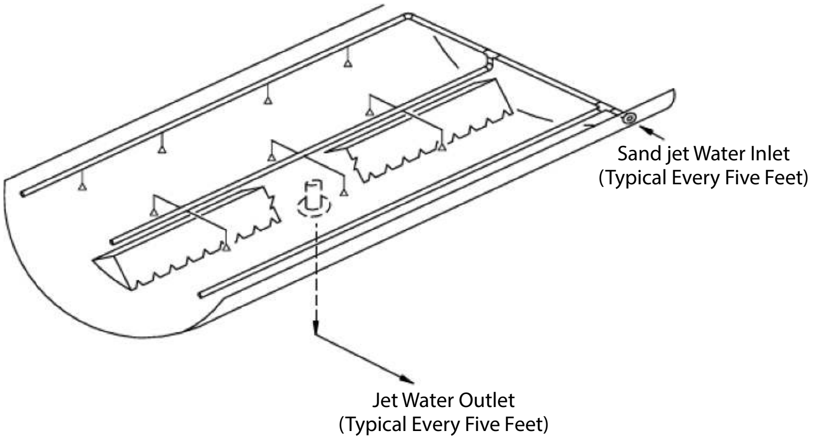


Figure 18.90 Schematic of a horizontal separator fitted with sand jets and inverted trough [79].

of sand can be removed by periodically injecting water or steam in the bottom of the vessel so as to suspend the sand during draining. Figure 18.90 illustrates a cutaway section of a sand wash and drain system fitted into a horizontal separator fitted with sand jets and an inverted trough. Removal of sand upstream of separators provides substantial operational benefits. Mozley Wellspin desander has been developed to remove sand effectively in simple, compact systems based on solid/liquid hydrocyclones, which remove the sand before it enters the separator. Sand problems may be solved by using a filter or desanding cyclone before the separator. But filters often block in sandy service and are not readily used.

Liquid Carryover

Liquid carryover occurs when free liquid escapes with the gas phase and can indicate high liquid level, thus causing damage to vessel internals, foam, improper design, plugged liquid outlets, or a flow rate that exceeds the design rate of the vessel. Liquid carryover can be prevented by installing a level safety high (LSH) sensor that shuts in the inlet flow to the separator when the liquid level exceeds the normal maximum liquid level by 10–15%.

Gas Blowby

Gas blowby occurs when free gas escapes with the liquid phase can be an indication of low liquid level, vortexing, or level control failure. This can result in a dangerous situation where there is a level control failure and the liquid dump valve is open; the gas entering the vessel will exit the liquid outlet line and would have to be handled by the next downstream vessel in the process. The downstream vessel needs to be designed for the gas blowdown condition otherwise it can over pressurize. It can be prevented by installing a level safety low sensor (LSL) that shuts the inflow and/or outflow to the vessel when the liquid level drops to 10–15% below the lowest operating level. Furthermore, the downstream process equipment is equipped with a pressure safety high (PSH) sensor and a pressure safety valve (PSV) sized for gas blowby.

Emulsion

Emulsions are often present problems in the operation of three-phase separators. Over a period of time an accumulation of emulsified materials and/or other impurities usually will form at the interface of the water and oil phases. In addition to adverse effects on the liquid level control, this accumulation will also decrease the effective oil or water retention time in the separator, with a resultant decrease in water–oil separation efficiency. The addition of chemicals and/or heat often minimizes this effect. Furthermore, lowering the settling time required for oil–water separation is accomplished by either the application of heat in the liquid section of the separator, or the addition of demulsifying chemicals [81].

Piping Requirements

Pipes that are connected to and from the process vessels must not interfere with the good working of the vessels. Therefore, the following guidelines should be observed.

- There should be no valves, pipe expansions, or contractions within 10 pipe diameters of the inlet nozzle.
- There should be no bends within 10 pipe diameters of the inlet nozzle except the following:
  - For knockout drums and demisters, a bend in the feed pipe is permitted if this is in a vertical plane through the axis of the feed nozzle.
  - For cyclones, a bend in the feed pipe is allowed if this is in a horizontal plane and the curvature is the same direction as the cyclone vortex.
- A pipe reducer may be used in the vapor line leading from the separator, but it should be no nearer to the top of the vessel than twice the outlet pipe diameter.
- A gate or ball type valve that is fully opened in normal operation should be used, where a valve in the feed line near the separator cannot be avoided.
- High pressure drops that cause flashing and atomization should be avoided in the feed pipe.
- If a pressure reducing valve in the feed pipe cannot be avoided, it should be located as far upstream of the vessel as practicable.

Cyclone Separators

Cyclones are widely used for the separation and recovery of industrial dusts from air or process gases. Cyclones are the principal type of gas–solids separators using centrifugal force. They are simple to construct, of low cost, and are made from a wide range of materials with an ability to operate at high temperatures and pressures. Cyclones are suitable for separating particles where agglomeration occurs. Pollution and emission regulations have compelled designers to study the efficiency of cyclones.

Cyclones offer the least expensive means of dust collection. They give low efficiency for collection of particles smaller than 5 μm. A high efficiency of 98% can be achieved on dusts with particle sizes of 0.1 to 0.2 μm that are highly flocculated.

Cyclone reactor types permit study of flow pattern and residence time distribution [38, 39]. For example, see the studies by Coker [40, 41] of synthetic detergent production with fast reaction. Reactor type cyclones are widely used in separating a cracking catalyst from vaporized reaction products.

Reversed flow cyclones are the most common design, in which the dust-laden gas stream enters the top section of the cylindrical body either tangentially or via an involute entry. The cylindrical body induces a spinning vortexed flow pattern to the gas–dust mixture. Centrifugal force separates the dust from gas stream; the dust travels to the walls of the cylinder, and down the conical section to the dust outlet. The spinning gas also travels down the wall toward the apex of the cone, but reverses direction in an air core and leaves the cyclone through the gas outlet tube at the top. This consists of a cylindrical sleeve, the vortex finder, whose lower end extends below the level of the feed port. Separation depends on particle settling velocities, governed by size, density and shape.

Figure 18.91 gives a good typical cyclone arrangement, but this is by no means the highest efficiency or best design. Some researchers [42–44] have provided good design and performance analysis. Straimand [45] and Strauss [46] have given guidelines for designing cyclones. The effects of feed and cyclone parameters on the efficiency are complex, as many parameters are interdependent. Figures 18.92–18.94 show photographs of some commercial units and Table 18.29 gives the effects of important operating and design parameters on cyclone performance.

The zone of most efficient separation is in the conical region designated by dimension Z_c in Figure 18.91. The larger particles are thrown against the wall before the outlet was reached, the finer particles are thrown out in the inner vortex as the direction of motion is reversed. Here, the relative velocity difference between the particle and the carrier is the greatest for any point in the separator. Although the tangential velocity component predominates throughout the cyclone, the axial velocity prevails in the turbulent center. Van Dongen and ter Linden [47] measured pressure patterns in a typical cyclone and found the lowest total pressure at the extreme bottom point of the cone, even lower

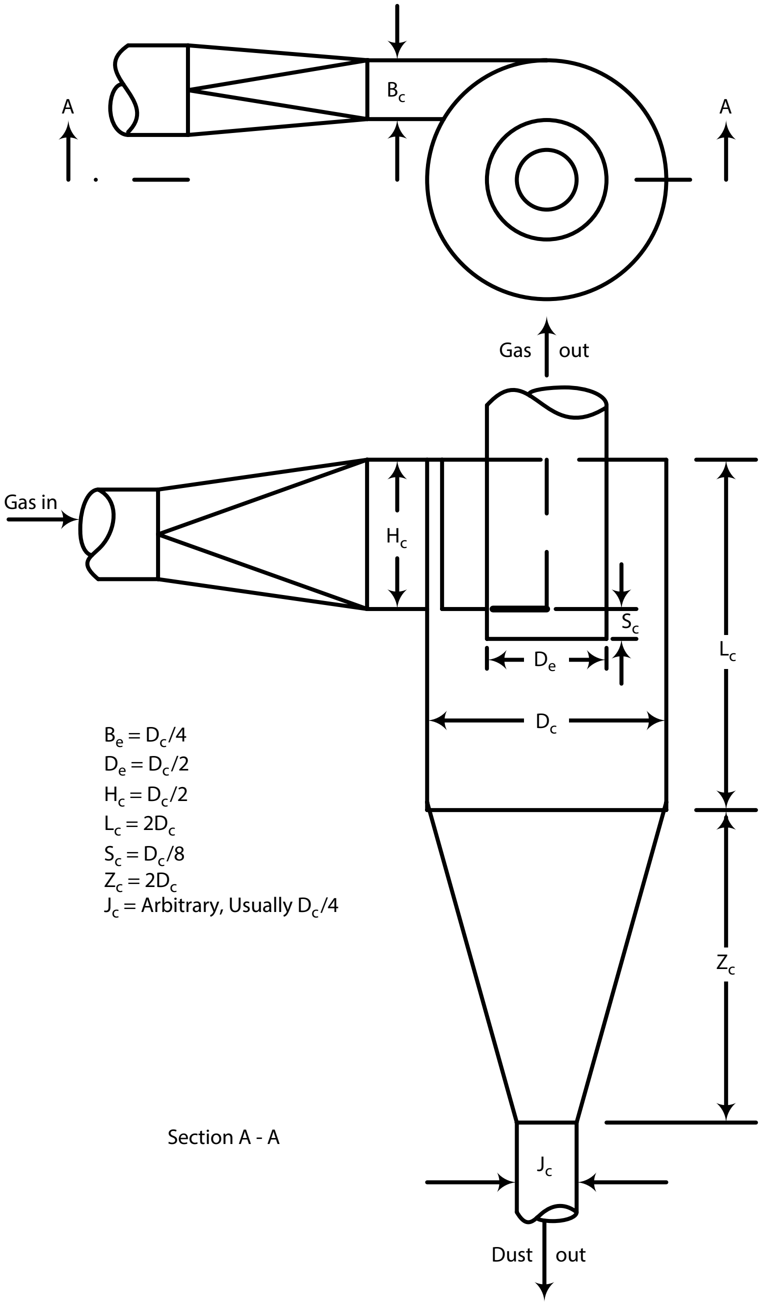


Figure 18.91 (a) Cyclone-separator proportions-dust system (by permission from Perry, J. H., Chemical Engineers Handbook, 3rd ed., McGraw-Hill Company, 1950). (*Continued*)

than at the gas exhaust. Their pressure profiles indicate considerable eddy or secondary gas movement in the unit near the vertical axis.

Solid Particle Cyclone Design

Following the general dimensional relations of the typical cyclone as shown in Figure 18.91, the following general guides apply. This cyclone is better suited to solid particles removal than liquid droplets. To avoid re-entrainment, it is important to keep the separated material from entering the center vortex of the unit. Solid particles generally slide down the walls with sufficient vertical velocity to avoid re-entrainment.

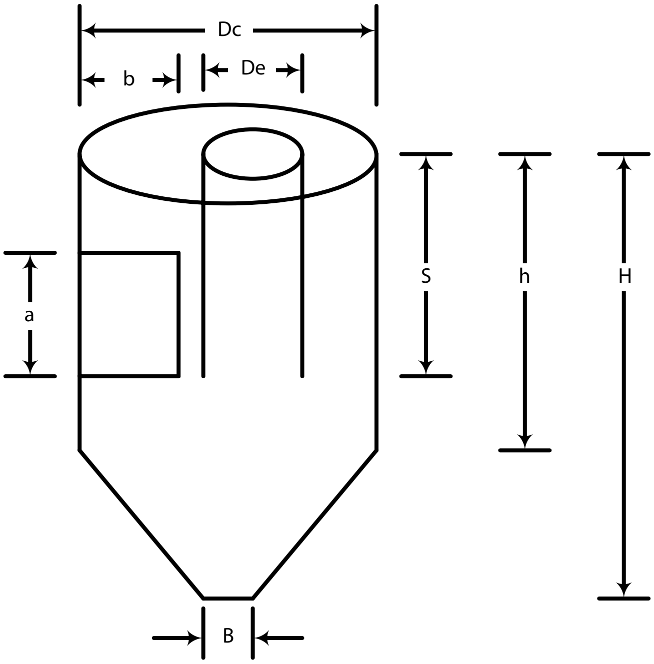


Figure 18.91 (Continued) (b) A cyclone: Design configurations.

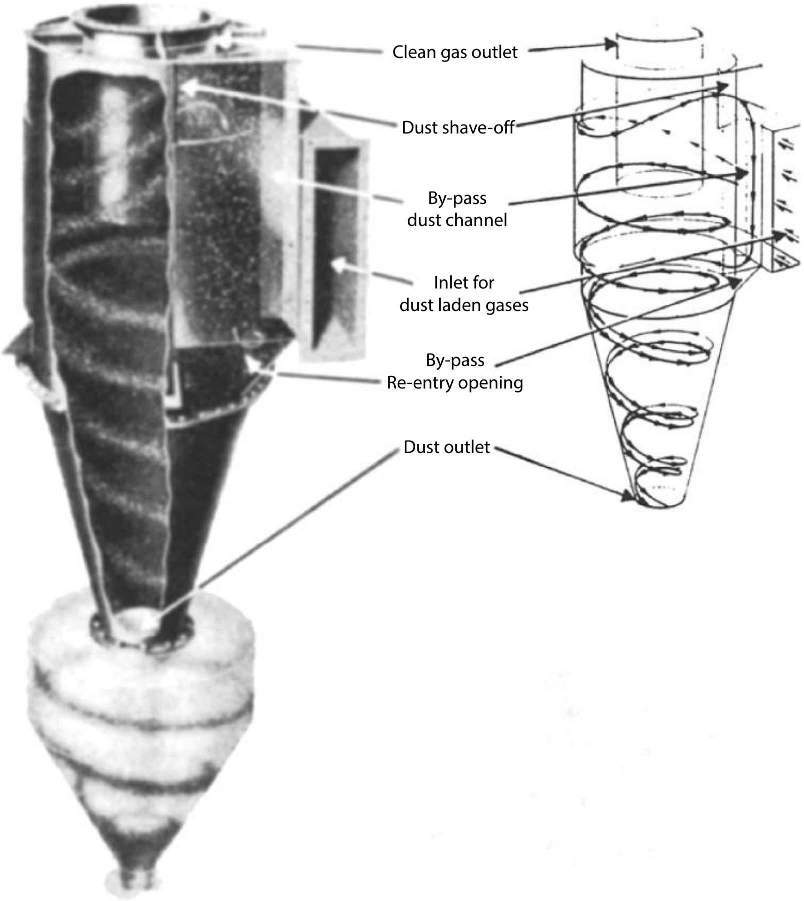


Figure 18.92 Van Tongeran dust shave-off design (courtesy of Buell Engineering Co.).

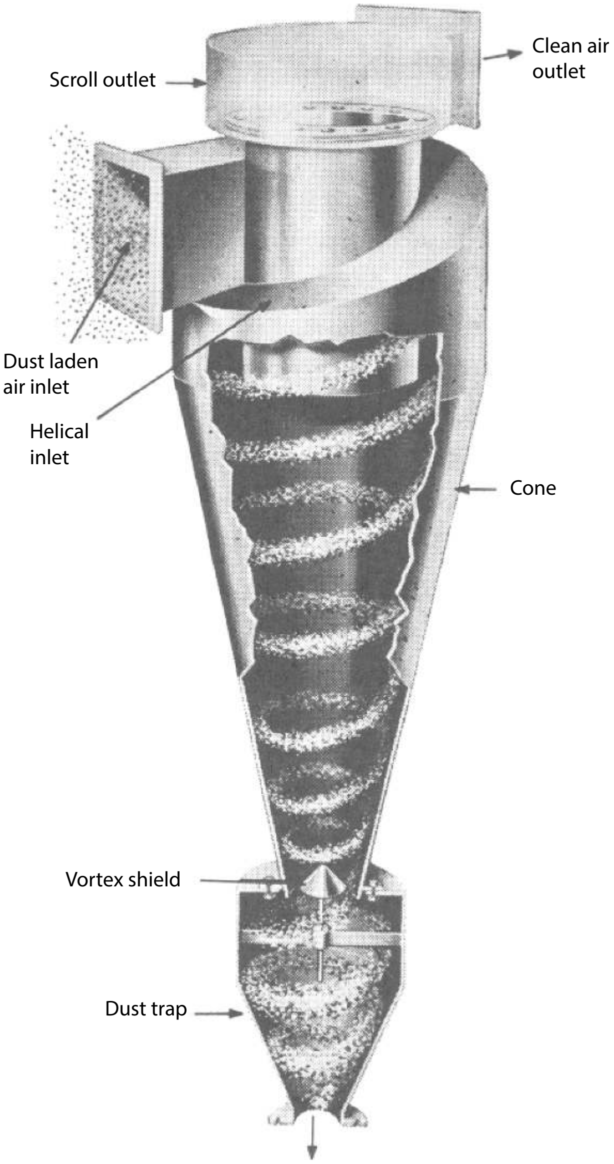


Figure 18.93 Helical entry cyclone (courtesy of the Ducon Co. Inc.).

a. Select outlet diameter to give gas velocity out not exceeding 600 ft/min. Bear in mind that higher velocity can be used in special designs.

b. Due to the usual conditions of limiting pressure drop, entrance velocities range from 1000 to 4000 ft/min. A velocity of 3000 ft/min is good average, although velocities to 6000 ft/min are used in some applications.

c. Select cylindrical shell diameter, D_c, with two considerations in mind:
- Large diameter reduces pressure drop
- Small diameter has higher collection efficiency for the same entrance conditions and pressure drop.

d. The length of the inverted cone section, Z_c, is critical, although there is no uniformity in actual practice. The dimensions suggested in Figure 18.91 are average.

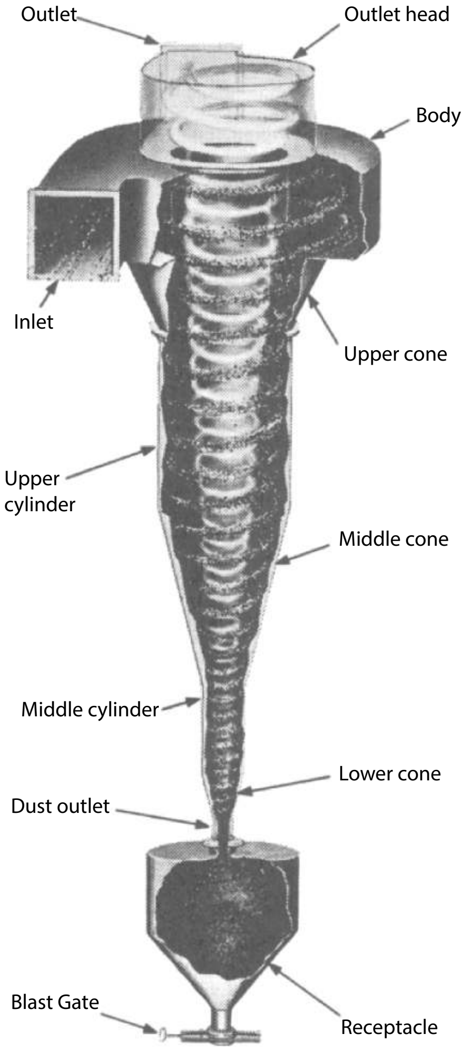


Figure 18.94 Involute entry cyclone (by permission from American Blower Div. American Radiator and Standard Sanitary Corp.).

Table 18.29 Effects of variables on cyclone performance.

Variable	Effect
Pressure drop increases	Cut size (diameter on particles of which 50% are collected) decreases, flow rate increases; sharpness increases
Solids content of feed increases	Cut size increases (large effect above 15–20% v/v)
$(\rho_p - \rho_f)$ increases	Cut size decreases
Viscosity of liquid phase increases	Little effect below 10 mPas
Cyclone diameter (D_c) increases	Cut size increases; pressure drop usually decreases
Cyclone inlet (a) diameter increases	Gravitational force in cyclone decreases; cut size increases; capacity falls; pressure drop decreases
Overflow diameter increases	Cut size increases; risk of coarse size appearing
Underflow diameter increases	Bring excess fines from liquid phase into underflow
Cyclone shape becomes longer	Decreases cut size; sharpens separation

Cyclone Design Procedure

The computation of cyclone fractional grade efficiency depends on cyclone parameters and flow characteristics of particle laden gases. The procedure involves a series of equations containing exponential and logarithmic functions. Koch and Licht [42] describe a cyclone using seven geometric ratios in terms of its diameter as:

$$\frac{a}{D_c}, \frac{b}{D_c}, \frac{D_e}{D_c}, \frac{S}{D_c}, \frac{h}{D_c}, \frac{H}{D_c}, \frac{B}{D_c}$$

They further stated that certain constraints are observed in achieving a sound design. They are as follows:

1. $a < S$ to prevent short-circuiting;
2. $b < \frac{1}{2}(D_c - D_e)$ to avoid sudden contraction;
3. $S + 1 \leq H$ to keep the vortex inside the cyclone
4. $S < h$
5. $h < H$
6. $\Delta P < 10$ in. H_2O
7. $v_i / v_s \leq 1.35$ to prevent re-entrainment

where
v_i = inlet linear velocity, ft/s
v_s = Saltation velocity, ft/s

8. $v_i / v_s = 1.25$ for optimum efficiency.

The Equations

1. Natural length, l: is the distance below the gas outlet where the vortex turns (Figure 18.91B).

$$l = 2.3 D_e \left(\frac{D_c^2}{ab} \right)^{1/3} \qquad (18.275)$$

For l < (H –S), the cyclone volume at the natural length (excluding the core) is V_{nl}:

$$V_{nl} = \frac{\pi D_c^2}{4}(h - S) + \left(\frac{\pi D_c^2}{4} \right)\left(\frac{l + S - h}{3} \right)\left(1 + \frac{d}{D_c} + \frac{d^2}{D_c^2} \right) - \frac{\pi D_e^2 l}{4} \qquad (18.276)$$

The diameter of a central core at a point of vortex turns, d is:

$$d = D_c - (D_c - B)\left(\frac{S + l - h}{H - h} \right) \qquad (18.277)$$

For l > (H- S), the cyclone volume below the exit duct (excluding the core), V_H is

$$V_H = \frac{\pi D_c^2}{4}(h - S) + \left(\frac{\pi D_c^2}{4} \right)\left(\frac{H - h}{3} \right)\left(1 + \frac{B}{D_c} + \frac{B^2}{D_c^2} \right) - \frac{\pi D_c^2}{4}(H - S) \qquad (18.278)$$

For a vortex exponent n:

$$n = 1 - \left\{1 - \frac{(12D_c)^{0.14}}{2.5}\right\}\left\{\frac{T + 460}{530}\right\}^{0.3} \tag{18.279}$$

The cyclone volume constant K_c using V_{nl} or V_H

$$K_c = \frac{(2V_s + V_{nl,H})}{2D_c^3} \tag{18.280}$$

where

$$V_s = \left\{\frac{\pi(S - a/2)\left(D_c^2 - D_e^2\right)}{4}\right\} \tag{18.281}$$

The relaxation time, τ_i for particle species i of diameter d_{pi} is

$$\tau_i = \frac{\rho_p (d_{pi})^2}{18\mu}, s \tag{18.282}$$

Cyclone configuration factor G is specified by the geometric ratios that describe the cyclone's shape. The cyclone configuration factor G is a function only of the configuration and is specified by the seven geometrical ratios that describe its shape. G is expressed as:

$$G = \frac{8K_c}{K_a^2 K_b^2} \tag{18.283}$$

where

$$K_a = \frac{a}{D_c}, K_b = \frac{b}{D_c} \tag{18.284}$$

Substituting the values of K_a, K_b and K_c in Eq. 18.283 gives

$$G = \left\{2\left[\pi(S - a/2)\left(D_c^2 - D_e^2\right)\right] + 4V_{H,nl}\right\}\frac{D_c}{a^2 b^2} \tag{18.285}$$

The fractional or grade efficiency η_i can be expressed as:

$$\eta_i = 1 - \exp\left\{-2\left[\frac{G\tau_i Q}{D_c^3}(n+1)\right]^{0.5/(n+1)}\right\} \tag{18.286}$$

where

a = inlet height, ft
b = inlet width, ft
B = cyclone dust outlet diameter, ft
d = diameter of central core at a point where vortex turns, ft
d_{pi} = diameter of particle in size range, i, ft
D_e = cyclone gas exit duct diameter, ft
D_c = cyclone diameter, ft
$(D_p)_{crit}$ = critical particle diameter, ft
g = acceleration due to gravity, 32.2 ft/s^2
G = cyclone configuration factor
h = cylindrical configuration factor
H = height of a segment of a circle, ft
i = subscript denotes interval n particle size range
K_a = a/D_c
K_b = b/D_c
K_c = cyclone volume constant
l = natural length (distance below gas outlet where vortex turns), ft
n = vortex component
Q = total gas flow rate, actual ft^3/s
S = gas outlet length, ft
T = temperature, °F
V_H = volume below exit dust (excluding core), ft^3
V_{nl} = volume at natural length (excluding core), ft^3
V_s = annular volume above exit duct to middle of entrance duct, ft^3
η_i = grade efficiency for particle size at midpoint of internal, i %.
τ = relaxation time, s.

Saltation Velocity

Koch and Licht [42] expressed the saltation velocity as the minimum fluid velocity necessary to prevent the settling out of solid particles carried in the stream and as the necessary velocity that picks up deposited particles and transports them without settling.

Zenz [48] has shown that the velocity given by the latter differs from the former by a factor of 2–2.5. Kalen and Zen [49] have applied the saltation concept to cyclone design by assuming:

1. There is no slippage between fluid and particles. The cyclone inlet width is the effective pipe diameter for calculating saltation effects.
2. Grain loading (dust concentration) is less than 10 grains/ft^3.
3. The diameter effect on the saltation velocity is proportional to the 0.4 power of the inlet width.

The saltation velocity, v_s is dependent on cyclone dimensions, as well as particle and fluid properties; v_s is expressed as:

$$v_s = 2.055\omega\left[\frac{b/D_c}{\left(1-b/D_c\right)^{1/3}}\right]D_c^{0.067}\,v_i^{2/3},\ \text{ft/s} \tag{18.287}$$

where

$$\omega = \left\{\frac{4g\mu(\rho_p-\rho_f)}{3\rho_f^2}\right\}^{1/3} \tag{18.288}$$

Inlet velocity, v_i,

$$v_i = \frac{Q}{(ab)}, \text{ ft/s} \tag{18.289}$$

Kalen and Zenz have shown that maximum cyclone collection efficiency occurs at $v_i/v_s = 1.25$, and Zenz has found experimentally that fluid re-entrainment occurs $v_i/v_s = 1.36$.

Pressure Drop (ΔP)

Several attempts have been made to calculate the frictional loss or ΔP of a cyclone, although none has been very satisfactory. Assumptions made have not considered entrance compression, wall friction, and exit contraction, all of which have a major effect. Consequently, no general correlation of cyclone ΔP has been adopted. Pressure drop in a cyclone with collection efficiency is important in evaluating its cost. Correlations for the pressure drop have empirical values, and are acceptable up to ΔP = 10 in. H_2O. The pressure drop ΔP, or the frictional loss is expressed in terms of the velocity head based on the cyclone inlet area. The frictional loss through cyclones is from 1 to 20 inlet velocity heads, and depends on the geometric ratios; ΔP through a cyclone is given by

$$\Delta P = 0.003\rho_f v_i^2 N_H \tag{18.290}$$

where

$$N_H = K\left(\frac{ab}{D_e^2}\right) \tag{18.291}$$

and

N_H = number of inlet velocity heads
ΔP = pressure drop, in H_2O
v_i = inlet velocity, ft/s
ρ_f = fluid density, lb/ft^3

K = 16 for no inlet vane; K = 7.5 with a neutral inlet vane. ΔP depends strongly on the inlet velocity, and high velocities can cause both re-entrainment and high pressure drop. However, entrainment can be reduced to a minimum, if the cyclone has a small base angle.

Figure 18.95 shows stationary vane centrifugal separator.

Critical Particle Diameter

Many theories have been proposed to predict the performance of a cyclone, although no fundamental relationship has been accepted. Capacity and efficiency depend on the inlet velocity and the dimensions of the vessel. Attempts have been made to predict the critical particle diameter, $(D_p)_{crit}$. This is the size of the smallest particle that is theoretically separated from the gas stream with 50% efficiency. The critical particle diameter is defined by Walas [50] as

$$(D_p)_{crit} = \left[\frac{9\mu D_c}{4\pi N_t v_i(\rho_p - \rho_g)}\right]^{0.5} \tag{18.292}$$

where

$$N_t = (v_i)\left[0.1079 - 0.00077v_i + 1.924(10^{-6})v_i^2\right] \tag{18.293}$$

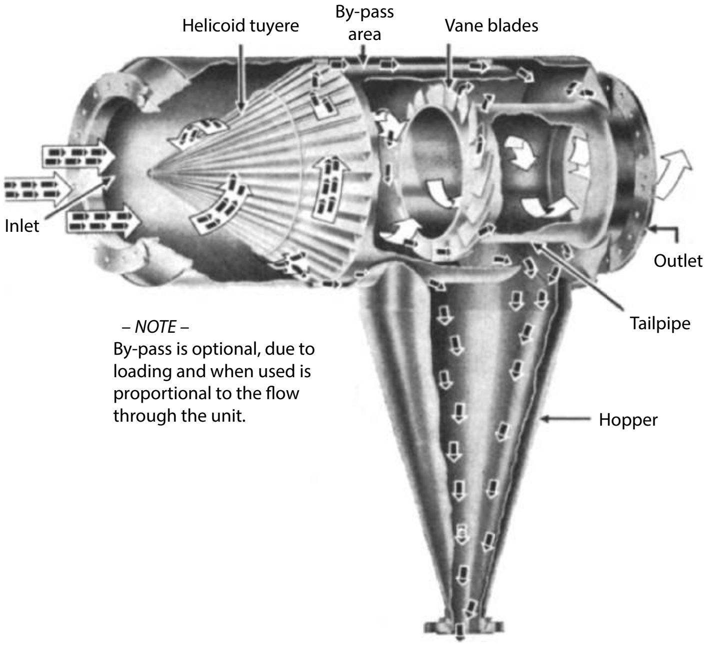


Figure 18.95 Stationary vane centrifugal separator (courtesy of Centrifix Corp.).

and

D_c = cyclone diameter, ft
v_i = the inlet linear velocity, ft/s
μ = fluid viscosity, lb/ft s
ρ_f = fluid density, lb/ft^3
ρ_p = particle density, lb/ft^3
N_t = effective number of turns made by the gas stream in the cyclone.

N_t has been found to be about 5 turns of the gas stream in the unit, and is considered somewhat conservative. When re-entrainment takes place, N_t may drop to 1.0 or 2.0. The API study presents an excellent survey of cyclone dust collectors [51].

With a height of opening equal to 2.5 times the width, the volumetric rate is

$$Q = Av_i = 2.5D_c\, v_i / 16 \tag{18.294}$$

To obtain a high efficiency, the vessel diameter must be small, but in order to accommodate a required volumetric rate, many units in parallel may be needed. These units, called multi-cyclones, may be incorporated in a single shell at a cost that may be justifiable in view of greater efficiency and lower pressure drop. An important parameter is the critical particle diameter, which is one that is removed to the extent of 50%. Figure 18.96 shows the percentage removal of particles in a cyclone as a function of the ratio of the particle diameter to the critical diameter.

Figure 18.97 shows the general efficiency curves of helical entry cyclone dusts separators and Figure 18.98 illustrates the general efficiency curves to involute cyclone dusts separator respectively.

Cyclone Design Factors

Cyclones are designed to meet specified ΔP limitations. The factor that controls the collection efficiency is the cyclone diameter, and a smaller diameter cyclone at a fixed ΔP will have a higher efficiency. Therefore, small diameter cyclones require a multiple of units in parallel for a given capacity. Reducing the gas outlet diameter results in an increased collection efficiency and ΔP. High efficiency cyclones have cone lengths in the range of 1.6 to 3.0 times the cyclone diameters.

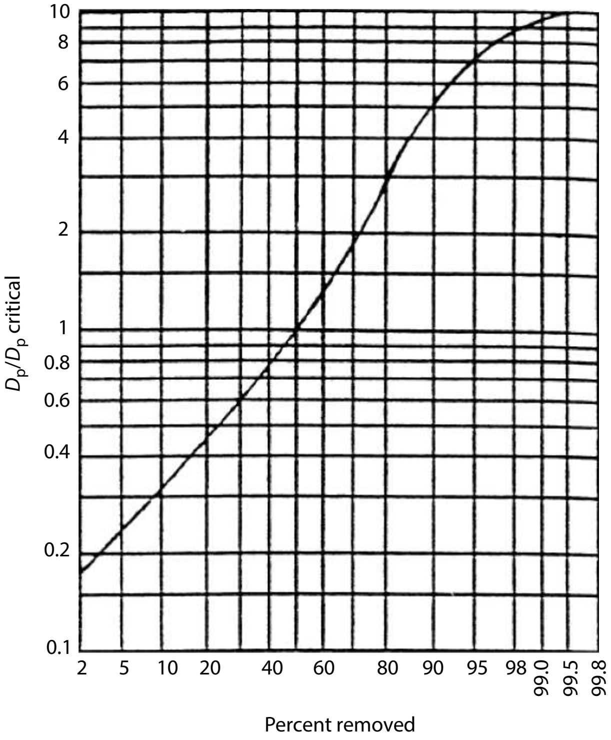


Figure 18.96 Percentage removal of particles as a function of particle diameter relative to the critical diameter.

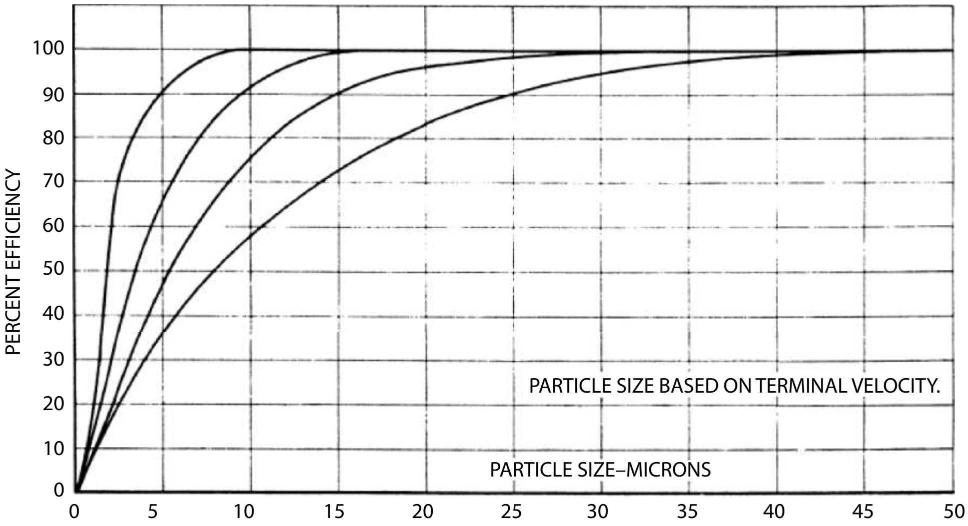


Figure 18.97 General efficiency curves, applied specifically to helical entry cyclone dust separators (courtesy of the Ducon Co.).

Collection efficiency increases as the gas throughput increases. Kalen and Zenz [49] reported that collection efficiency increases with increasing gas inlet velocity to a minimum tangential velocity. This reaches the point where the dust is re-entrained or not deposited because of saltation. Koch and Licht [42] showed that saltation velocity is consistent with cyclone inlet velocities in the range of 50 to 90 ft/s.

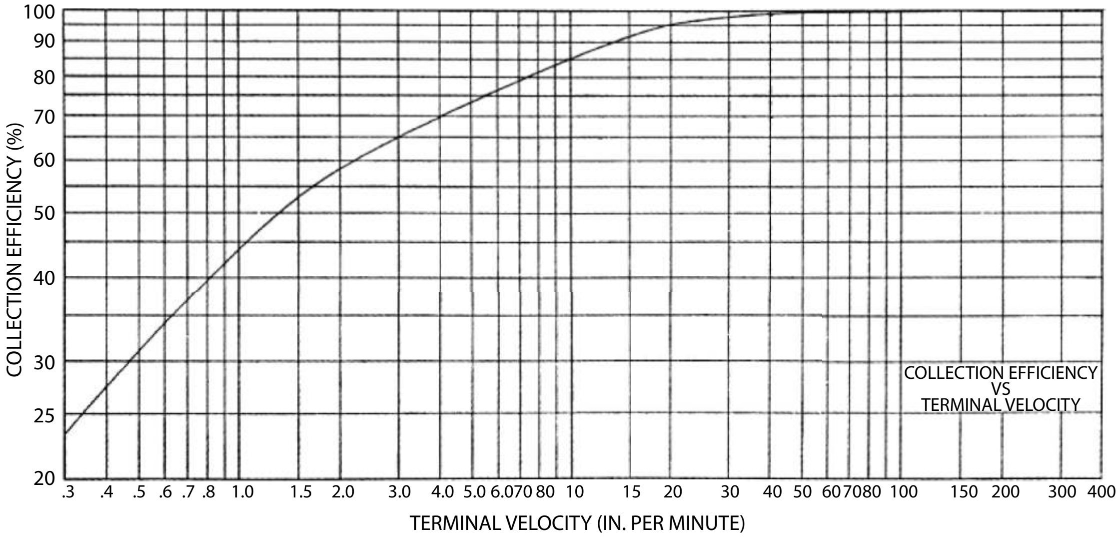


Figure 18.98 General efficiency curve applies specifically to involute entry cyclone dust separators (courtesy of American Blower Div., American Radiator and Standard Sanitary Corp.).

Troubleshooting Cyclone Maloperations

In general, cyclones are used to separate particles from the gas stream, but recent developments have enabled cyclones to function as reactors. Some cyclone reactors can separate cracking catalyst from vaporized reaction products in the range of 950°F and 1000°F, or can function as regenerators for flue gases between 1250°F and 1500°F. In both cases, the high particles velocities can cause rapid erosion of the cyclone material. This often results in poor performance of the cyclone. Other causes of poor cyclone performance are:

- a hole in a cyclone body;
- a cyclone volute plugged;
- a dip leg unsealed;
- a dip leg plugged and
- dip leg failure.

Lieberman [52] has reviewed the causes of these maloperations, which often result in loss catalyst and reduced efficiency. A deficient cyclone reactor is identified by bottom sediment and water levels in the slurry oil products. For a regenerator cyclone, problems are visibly identified by the increased opacity of the regenerator flue gas or by reduced rates of spent catalyst withdrawal.

Example 18.19

Determine the efficiency and pressure drop (ΔP) based on cyclone dimensions and the gas flow rate at 516.7 ft^3/sand density of 0.075 lb/ft^3 containing particles with density of 62.43 lb/ft^3. The dimensions and data required are shown in the following Table

Data	**Dimension**
Cyclone gas inlet height, ft.	4.5
Cyclone gas inlet width, ft	1.896
Cyclone gas outlet length, ft	3.448
Cyclone gas outlet diameter, ft	3.792

Cyclone cylindrical height, ft	8.552
Cyclone diameter, ft	6.333
Gas rate, ft^3/s	516.7
Gas density, lb/ft^3	0.075
Particle density, lb/ft^3	62.43
Gas viscosity, lb/ft.s	1.28×10^{-5}
Temperature, °F	110
Particle size, ft	3.281×10^{-5}

Solution

1. Natural length, l, is the distance below the gas outlet where the vortex turns. From Eq. 18.275

$$l = 2.3 D_e \left(\frac{D_c^2}{ab} \right)^{1/3}$$

$$l = 2.3(3.792) \left(\frac{6.33^2}{4.5 \times 1.896} \right)^{1/3} = 14.597\,\text{ft}$$

For l < (H –S), the cyclone volume at the natural length (excluding the core) is V_{nl}:

Check if l < (H-S)

$$14.597 < (26.333 - 3.448)$$

$$14.597 < 22.885$$

Cylindrical volume at the natural length, V_{nl} is calculated using Eq. 18.276.

$$V_{nl} = \frac{\pi D_c^2}{4}(h - S) + \left(\frac{\pi D_c^2}{4} \right) \left(\frac{l + S - h}{3} \right) \left(1 + \frac{d}{D_c} + \frac{d^2}{D_c} \right) - \frac{\pi D_e^2 l}{4}$$

$$V_{nl} = \frac{\pi (6.333)^2}{4}(8.552 - 3.448) + \left(\frac{\pi 6.333^2}{4} \right) \left(\frac{14.597 + 3.448 - 8.552}{3} \right) \left(1 + \frac{4.304}{6.333} + \frac{4.304^2}{6.333^2} \right)$$
$$- \frac{\pi (3.792)^2 (14.597)}{4}$$
$$= 209.37\,\text{ft}^3$$

The diameter d of a central core at a point of vortex turn is calculated from Eq. 18.277.

$$d = 6.333 - (6.333 - 2.533)\left[\frac{(3.448 + 14.597 - 8.552)}{(26.333 - 8.552)}\right] = 4.304 \text{ ft}$$

From Eq. 18.279, a vortex exponent n is

$$n = 1 - \left\{1 - \frac{(12 \times 6.333)^{0.14}}{2.5}\right\}\left\{\frac{(110 + 460)}{530}\right\}^{0.3} = 0.767$$

From Eq. 18.281

$$V_s = \left\{\frac{\pi(3.448 - 4.5/2)(6.333^2 - 3.792^2)}{4}\right\} = 24.207 \text{ ft}^3$$

Then, from Eq. 18.280

$$K_c = \frac{(2 \times 24.207 + 209.37)}{(2 \times 6.333^3)} = 0.507$$

The relaxation time, τ_i for particle species i of diameter d_{pi} (from Eq. 18.282) is

$$\tau_i = \frac{(62.43)(0.00003281)^2}{(18)(0.0000128)} = 0.0002916 \text{ s}$$

Cyclone configuration factor G (from Eq. 18.285) is

$$G = \{2[\pi(3.448 - 4.5/2)(6.333^2 - 3.792^2)] + 4(209.37)\}\frac{6.333}{(4.5^2)(1.896^2)}$$
$$= 89.71$$

Inlet velocity, v_i, ft/s (from Eq. 18.289) is

$$v_i = \frac{516.7}{(4.5)(1.896)} = 60.56 \text{ ft/s}$$

Saltation velocity, v_s (from Eq. 18.288) is

$$v_s = 2.055\omega\left[\frac{b/D_c}{\left(1 - b/D_c\right)^{1/3}}\right]D_c^{0.067}\, v_i^{2/3}, \text{ ft/s}$$

where

$$\omega = \left\{ \frac{4g\mu(\rho_p - \rho_f)}{3\rho_f^2} \right\}^{1/3}$$

$$\omega = \left[\frac{4 \times 32.2 \times 0.0000128(62.43 - 0.075)}{(3)(0.075)^2} \right]^{1/3} = 1.826$$

Then

$$v_s = (2.055)(1.826)\left[\frac{1.896/6.333}{(1 - 1.896/6.333)^{1/3}} \right](6.333)^{0.067} = 22.0775 \text{ ft/s}$$

The critical particle diameter (from Eq. 18.292) is

$$(D_p)_{crit} = \left[\frac{9\mu D_c}{4\pi N_t v_i(\rho_p - \rho_g)} \right]^{0.5}, \text{ft.}$$

where the effective number of turns, N_t (from Eq. 18.293) is

$$\begin{aligned} N_t &= (v_i)\left[0.1079 - 0.00077v_i + 1.924(10^{-6})v_i^2\right] \\ &= 60.56[0.1079 - 0.00077(60.56) + 1.924(10^{-6})(60.56)^2] = 4.137 \end{aligned}$$

The critical particle diameter from Eq. 18.292 is

$$\begin{aligned} (D_p)_{crit} &= \left[\frac{9(0.0000128)(6.333)}{4(\pi)(4.137)(60.56)(62.43 - 0.075)} \right]^{0.5} \\ &= 6.096 \times 10^{-5} \text{ ft } (19\,\mu\text{m}) \end{aligned}$$

The ratio of particle diameter to critical particle diameter is:

$$\frac{D_p}{(D_P)_{crit}} = \frac{0.00003281}{0.00006096} = 0.538$$

From Figure 18.104, 28% of the other sized particles are removed. The cyclone pressure drop ΔP from Eq. 18.290 is

$$\Delta P = 0.003\rho_f v_i^2 N_H, \text{in } H_2O$$

where (according to Eq. 18.293)

$$N_H = K\left(\frac{ab}{D_e^2}\right)$$

K = 16 for no inlet vane; K = 7.5 with a neutral inlet vane.

$$N_H = 16\left(\frac{(4.5)(1.896)}{(3.792)^2}\right) = 9.4936$$

$$\Delta P = 0.003(0.075)(60.56)^2(9.4936) = 7.83\,\text{in.}\,H_2O$$

The fractional or grade efficiency η_i (from Eq. 18.286) is

$$\eta_i = 1 - \exp\left\{-2\left[\frac{(89.71)(0.0002916)(516.7)}{(6.333)^3}(1+0.767)\right]^{0.5/(1.767)}\right\} = 0.641$$

The Excel spreadsheet (Example 18.19.xlsx) shows the calculations of cyclone sizing and the table below shows the results of Example 18.19.

Results of Cyclone Design			
Cyclone overall height, ft	H	26.333	ft.
Cylindrical height of cyclone, ft	h	8.552	ft.
Cyclone diameter, ft	Dc	6.333	ft.
Cyclone dust outlet diameter, ft.	B	2.533	ft.
Cyclone gas inlet length, ft.	a	4.5	ft.
Cyclone gas inlet width, ft	b	1.896	ft.
Cyclone gas outlet length, ft	S	3.448	ft.
Cyclone gas outlet diameter, ft	De	3.792	ft.
Gas rate, ft^3/s	Q	516.7	ft^3/s
Particle density, lb/ft^3	RHOp	62.43	lb/ft^3
Gas density, lb/ft^3	RHOg	0.075	lb/ft^3
Gas viscosity, lb/ft-sec.	VIS	0.0000128	lb/ft-s
Gas temperature, °F	T	110	°F
Particle size, ft.	Dp	0.00003281	ft
Vortex Exponent		0.728	
Relaxation time		0.000291691	s
Inlet fluid velocity		60.56	ft/s
Saltation velocity		22.078	ft/s
Cyclone configuration factor		89.728	
Cyclone pressure drop		7.834	in. H_2O
Ratio of the inlet to the Saltation velocities		2.74	

Cyclone efficiency	63.31	%
Effective number of turns	4.138	
Critical particle diameter	6.0954E-05	ft
Critical particle diameter	1.853E-05	micron
Ratio of particle diameter to critical diameter	0.538	

Cyclone Collection Efficiency

Typical estimating efficiencies are given in Figures 18.105 and 18.106. Note the curves indicate how much dust of each particle size will be collected. The efficiency increases as the pressure drop increases; that is, a smaller separator might have a higher efficiency due to the higher gas velocities and increased resistance than a larger unit for the same gas flow. For example, there are several curves of the typical shape of Figure 18.105, with each curve for a definite resistance to flow through the unit.

The pressure drop in a typical cyclone is usually between 0.5 and 8 in. of H_2O. It can be larger, but rarely exceeds 10 in. H_2O for single units. The API study [51] summarizes the various factors. Lapple [5, 53] gives calculation equations, but in general the most reliable pressure drop information is obtained from the manufacturer. Here is how the pressure drop maybe estimated.

For the typical cyclone of Figure 18.91 [5]

(a) Inlet velocity head based on inlet area:

$$h_{ui} = 0.003\rho V_c^2 \tag{18.295}$$

(b) Internal cyclone friction loss:

$$F_{cv} = \Delta P_{cv} + 1 - \left(\frac{4A_c}{\pi D_e^2}\right)^2 \tag{18.296}$$

where A_c = cyclone inlet area.

$$\Delta P_{cv} = k\left(\frac{D_c}{D_e}\right)^2 \tag{18.297}$$

where

D_c = cyclone diameter, ft

D_e = cyclone gas exit duct diameter, ft.

K has been found to be constant at 3.2 for cyclones with an involute entrance

$$B_e/D_c = 1/8 \text{ to } 3/8$$

$$H_c/D_c = 1.0$$

$$D_e/D_c = 1/4 \text{ to } 3/4$$

For the typical cyclone of Figure 18.91 with tangential inlet:

$$F_{cv} = \frac{KB_eH_c}{D_e^2} \tag{18.298}$$

where B_eH_c = cyclone inlet area, ft^2.

$$K = 16$$
$$F_{cv} = 8.0$$

If inlet vane is formed with inlet connection, K = 7.5.

Example 18.20 Cyclone System Pressure Drop

A cyclone system is to be installed as a part of a bagging operation. The unit is shown in Figure 18.99. Determine the head required for purchase of the fan. The conditions are:

Air volume: 4000 ft^3/min of air at 70°F

Air density: 0.075 lb/ft^3

Areas	**(ft^2)**	**Gas velocity (ft/s)**	**Velocity head* (in. H_2O)**
Inlet duct:	1.398	47.7	0.50
Cyclone inlet:	2.0	33.33	0.25
Cyclone exit duct:	3.14	21.2	0.10

*Velocity head, inches $H_2O = V_D^2/(16)(106)$, V_D = ft/min.

Friction Loss ① to ②:

$$N_{Re} = 398{,}000$$

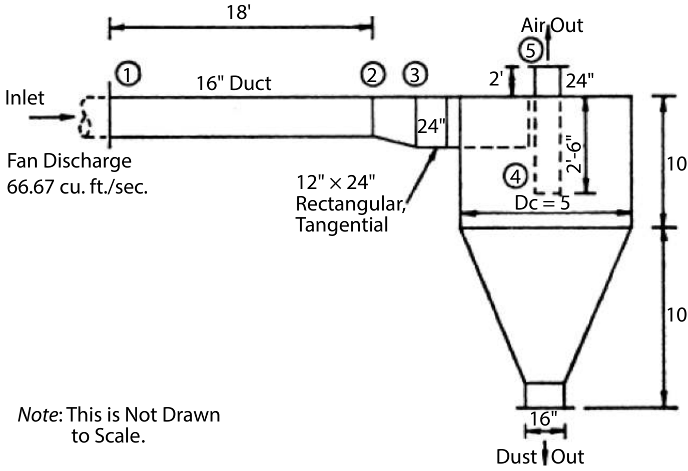


Figure 18.99 Pressure drop for cyclone separator system (adapted by permission from Lapple, C. E., Fluid and Particle Dynamics, 1st ed., Univ. of Delaware, 1954).

$$f = 0.0038$$

$$\text{No. Vel. Hd.} = \frac{4fL}{D} = \frac{4(0.0038)(18)}{(16/12)} = 0.204 \text{ vel. head}$$

$$\text{Loss} = (0.204)\ (0.5) = 0.102 \text{ in. } H_2O$$

Friction Loss ② to ③:

Assume as 1 Vel. head (conservative)

Friction loss = (1) (0.50) = 0.50 in. H_2O

Friction Loss ③ to ④:

$$F_{cv} = \frac{KW_iH_c}{D_e^2} = \frac{(16)(1)(2)}{(2)^2} = 8.0 \text{ Vel. head}$$

where W_i = width of rectangular cone inlet duct, ft.

Friction loss = 8.0 (0.25) = 2.0 in. H_2O

Friction Loss ④ to ⑤:

$$N_{Re} = 280{,}000$$
$$f = 0.004$$
$$\text{No. Vel. } H_{d.} = \frac{4fL}{D} = \frac{4(0.004)(4.5)}{(2)} = 0.036 \text{ Vel. head}$$

Friction loss = 0.036 (0.1) = 0.0036 in. H_2O

Since the unit exhausts to atmosphere with no additional restrictions, the total pressure drop is

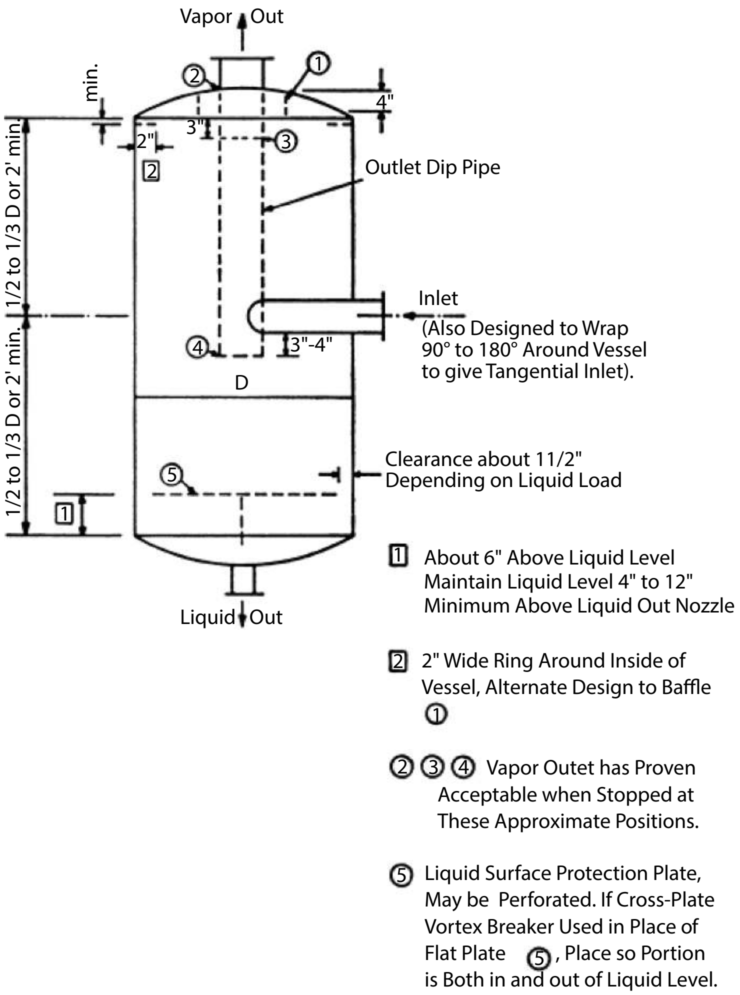
$\Delta P_{(total)}$ = Friction loss + downstream Vel. Head at ⑤—upstream vel. Head
$\Delta P_{(total)} = (0.102 + 0.50 + 2.0 + 0.0036) + 0^* - 0.50$
$\Delta P_{(total)} = 2.6056 + 0 - 0.50 = 2.10$ in. H_2O

Liquid Cyclone-Type Separator

The unit shown in Figure 18.100 has been used in many process applications with a variety of modifications [47, 54, 55]. It is effective in liquid entrainment separation, but is not recommended for solid particles due to the

---

* Note that point ⑤ is at atmospheric pressure and the velocity head is zero; however, if there had been a back pressure or resistance at this point before discharging it would have to be added in.



1 About 6" Above Liquid Level
Maintain Liquid Level 4" to 12"
Minimum Above Liquid Out Nozzle

2 2" Wide Ring Around Inside of Vessel, Alternate Design to Baffle ①

② ③ ④ Vapor Outet has Proven Acceptable when Stopped at These Approximate Positions.

⑤ Liquid Surface Protection Plate, May be Perforated. If Cross-Plate Vortex Breaker Used in Place of Flat Plate ⑤, Place so Portion is Both in and out of Liquid Level.

Figure 18.100 Centrifugal liquid separator.

arrangement of the bottom and outlet. The flat bottom plate serves as a protection to the developing liquid surface below. This prevents re-entrainment. In place of the plate a vortex breaker type using vertical cross plates of 4–12-in. depth also is used, (also see [56]). The inlet gas connection is placed above the outlet dip pipe by maintaining dimension of only a few inches at point 4. In this type unit some liquid will creep up the walls as the inlet velocity increases.

In order to handle higher loads, the liquid baffle is placed at the top to collect liquid and cause it to drop back down through the gas body. If the baffle is omitted, the liquid will run down the outlet pipe and be swept into the outlet nozzle by the outgoing gas as shown in Figure 18.101B. Figures 18.101 and 18.102 show several alternate inlet and outlet details. The unit with a tangential entry is 30–60% more efficient than one with only a turned-down 90° elbow in the center.

If the design of Figure 18.91 is used for liquid–vapor separation at moderately high liquid loads, the liquid sliding down the walls in sheets and ripples has somewhat of a tendency to be torn off from the rotating liquid and become re-entrained in the upward gas movement.

Liquid Cyclone Design (Based on Air–Water at Atmospheric Pressure)

For maximum liquid in outlet vapor of 4% wt based on incoming liquid to separator (see Figure 18.100).

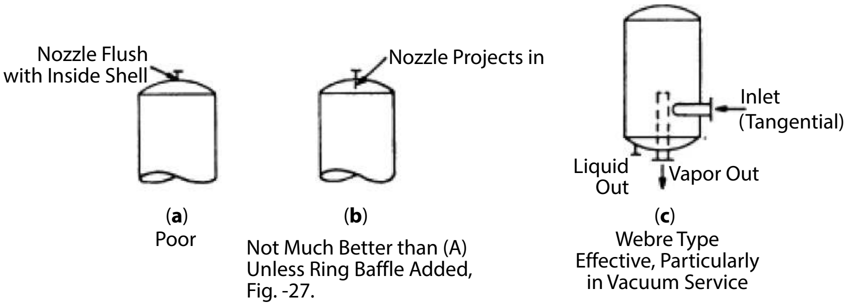


Figure 18.101 Separator outlets for liquid–vapor service.

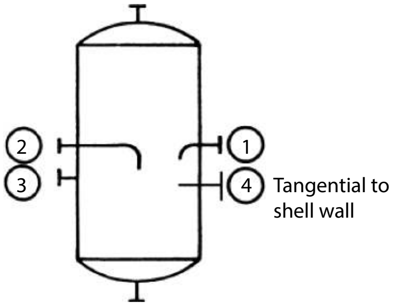


Figure 18.102 Separator inlets for liquid–vapor service.

- **a.** Select inlet pipe size to give vapor velocity at inlet of 100 to 400 ft/s for tangential pipe inlet.
- **b.** Select separator diameter to give velocity of 0.02 to 0.2 (max.) times the inlet velocity. At 400 ft/s pipe velocity, the separator velocity should be 0.018 to 0.03 times the pipe velocity. At 130 ft/s pipe velocity, the separator velocity should be 0.15 to 0.2 times the pipe velocity.
- **c.** Establish dimensions from typical unit of Figure 18.100. Always evaluate the expected performance in terms of the final design, adjusting vertical dimensions to avoid gas whipping on liquid films or droplets. Do not direct inlet gas toward an outlet. Place manway on same side of vessel as tangential inlet.
- **d.** Pressure drop is essentially negligible for the average conditions of use. Some estimate of entrance and exit losses can be made by fluid flow techniques, and an internal loss of 0.25 to 2.0 psi assumed, depending upon system pressure and general unit dimensions.
- **e.** For liquids and vapors other than air–water:

$$V_{(\text{separator})} = 0.1885 V_{sa}\left(\frac{\rho_L - \rho_v}{\rho_v}\right)^{0.25} \tag{18.299}$$

where V_{sa} is the selected separator velocity when using an air–water system, ft/s.

The Webre design as tested by Pollak and Work [57] showed (Figure 18.101c) that internal action in the separator was responsible for some of the entrainment, particularly liquid creep up the vessel walls.

The performance of this unit correlated for several different types of particle distribution by [57]:

$$\text{Log}\,[L_v(V'/L_l)^a] = b + cV' \tag{18.300}$$

a = 2 for Webre unit

L_v = entrainment, lb liquid/min/ft^2 of inlet

V′ = vapor velocity entering, lb/min/ft^2 of inlet
L_1 = liquid entering, lb/min/ft^2 of inlet

a, b, c, constants associated with the type and physical conditions of the system. For the unit of Pollak and Work [57]:

$$b = 1.85 \text{ and } c = 0.00643$$

Liquid–Solid Cyclone (Hydrocyclones) Separators

This type of solids removal device (Figures 18.103a and b) is a relatively low cost approach to remove/separate solids from solid/liquid suspensions. The incoming feed to such a unit is injected along the inner wall where the centrifugal force causes rotation at high angular velocity. The kinetic energy of these feeds is converted to centrifugal force. The coarse/heavier particles will be concentrated at the bottom as underflow. Most of the feed liquid and part of the fine solids will discharge through the vortex and overflow.

This unit is good to pre-thicken feed to centrifugal filters and similar applications. One cyclone may satisfy a requirement, or the units can be arranged in parallel for large capacities or in series for removal of extreme fines. Countercurrent wash system is given in Figure 18.104. Solids as small as 10 μm can be separated.

These units are made of abrasion resistant metals, solid plastics, or with corrosion-wear resistant plastic liners, such as molded rubber and elastomers; for example butyl, Hycar®, Hypalon®, urethane, and metal alloys, silicon carbide, alumina ceramics. These units require little or no maintenance.

The manufacturer requires complete solids and/or liquids data, feed size analysis, and requirements for separation. In some instances, it may be best to have a sample tested by the manufacturer in their laboratory.

Zanker, Reid and Voller [58, 59] give good performance analysis of these designs.

(**a**)

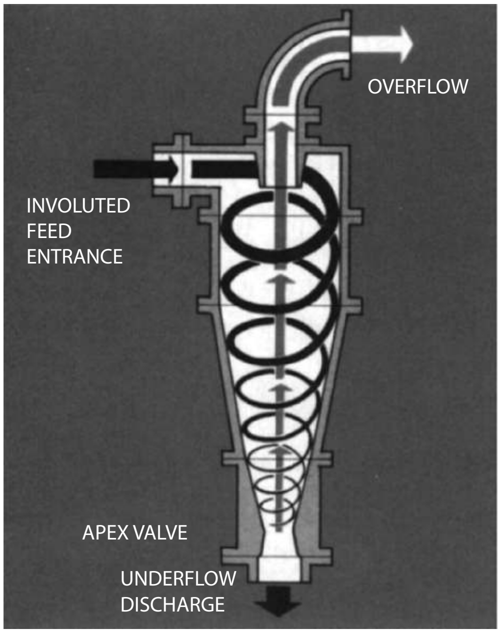
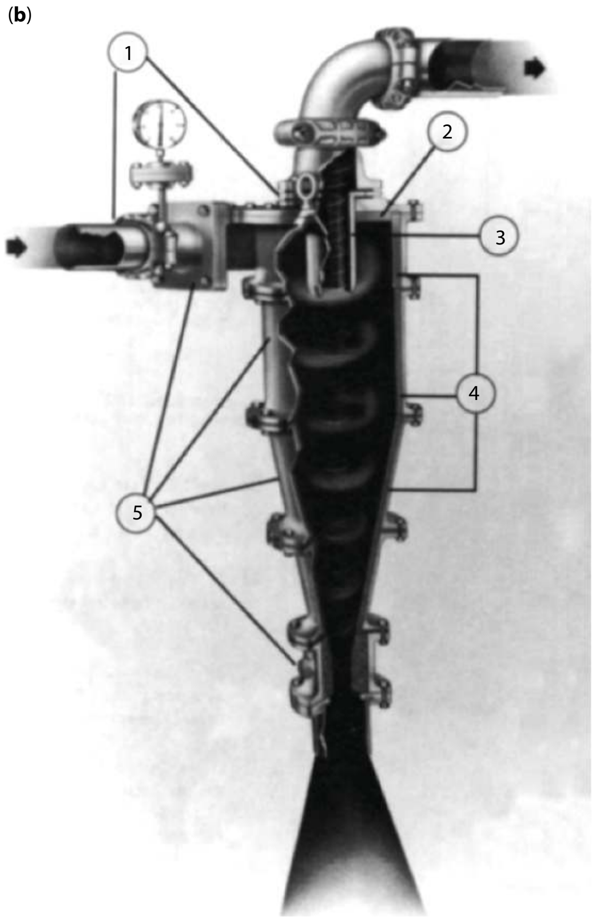


Figure 18.103a Liquid–solid removal cyclones. Feed enters tangentially along sidewall (by permission from Krebs Engineering).



1. Feed inlet and overflow connections are elastomer lined spool-piece adapters.
2. The top cover plate has all wetted surfaces lined, including the area mating with the overflow adapter.
3. The vortex finder is completely elastomer covered.
4. The molded liners for the inlet head, cylinder section, and conical sections have integral molded gaskets for sealing at the flanged joints. Molded liners and vulcanized linings are offered in gum rubber, polyurethane, nitrile rubber, butyl, Neoprene®, Viton®, Hypalon®, and other liner materials can be supplied.

Many of the molded elastomer liners are interchangeable with ceramic liners of silicon carbide or high purity alumina.

5. All metal housings are of cast or fabricated mild steel. Standard housings are for system pressures up to 25 psi, and special designs are available for higher system pressures.

Figure 18.103b Liquid–solid cyclone fabricated to resist corrosion and abrasion (by permission from Krebs Engineering).

Solid Particles in Gas/Vapor or Liquid Streams

The removal of solid particles from gas/vapor or liquid streams can be accomplished by several techniques, some handling the flow “dry,” others wetting the stream to settle/agglomerate the solids (or even dissolve) and remove the liquid phase from the system with the solid particles. Some techniques are more adaptable to certain industries

(**c**)

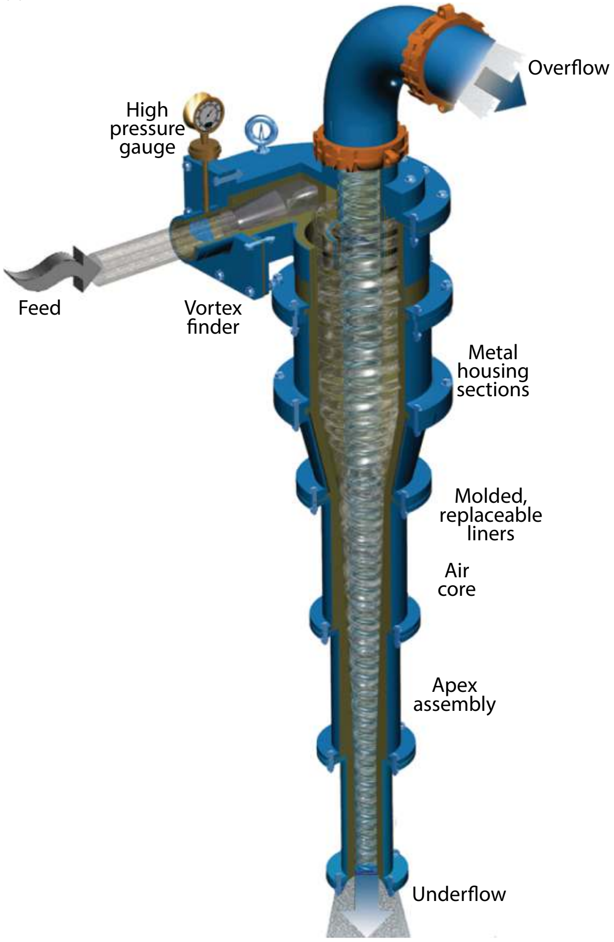


Figure 18.103c Photograph of a hydrocyclone (source: Krebs Engineers).

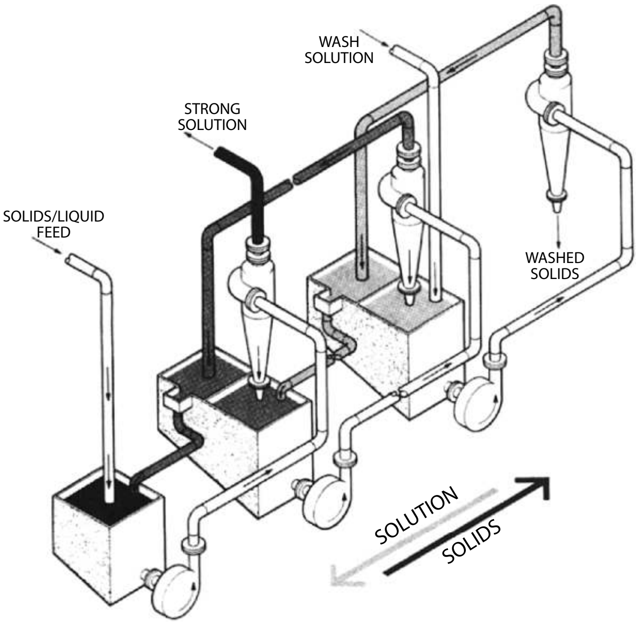


Figure 18.104 Cyclones used for countercurrent wash system (by permission from Krebs Engineers).

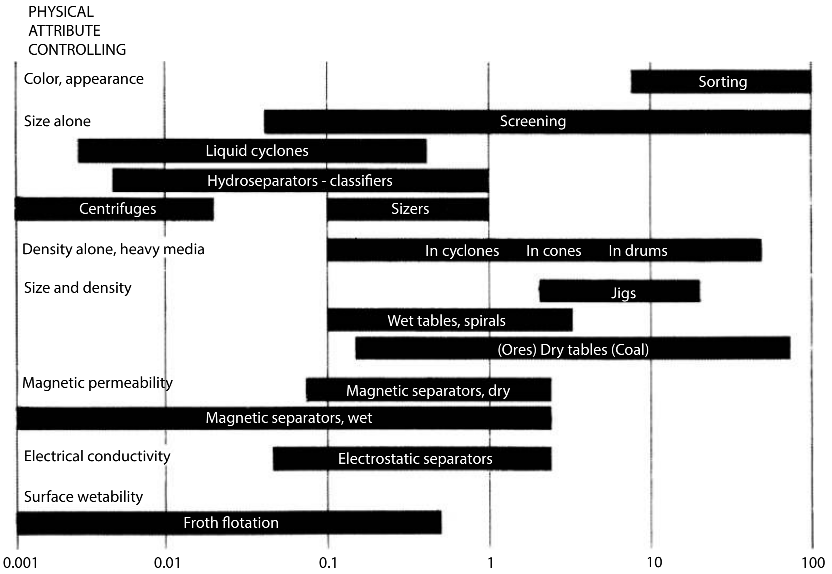


Figure 18.105 Size ranges where particular solid–solid/solid–liquid separation techniques can be applied (by permission from Roberts, E. J., *et al.* Chemical Engineering, 29 Jun. 1970 [31]).

than others. Figure 18.105 illustrates typical ranges of particle size removal of various types of common equipment or technique.

Inertial Centrifugal Separators

Specification Sheet, Figure 18.106, can be used as a guide in summarizing and specifying conditions for this type of equipment.

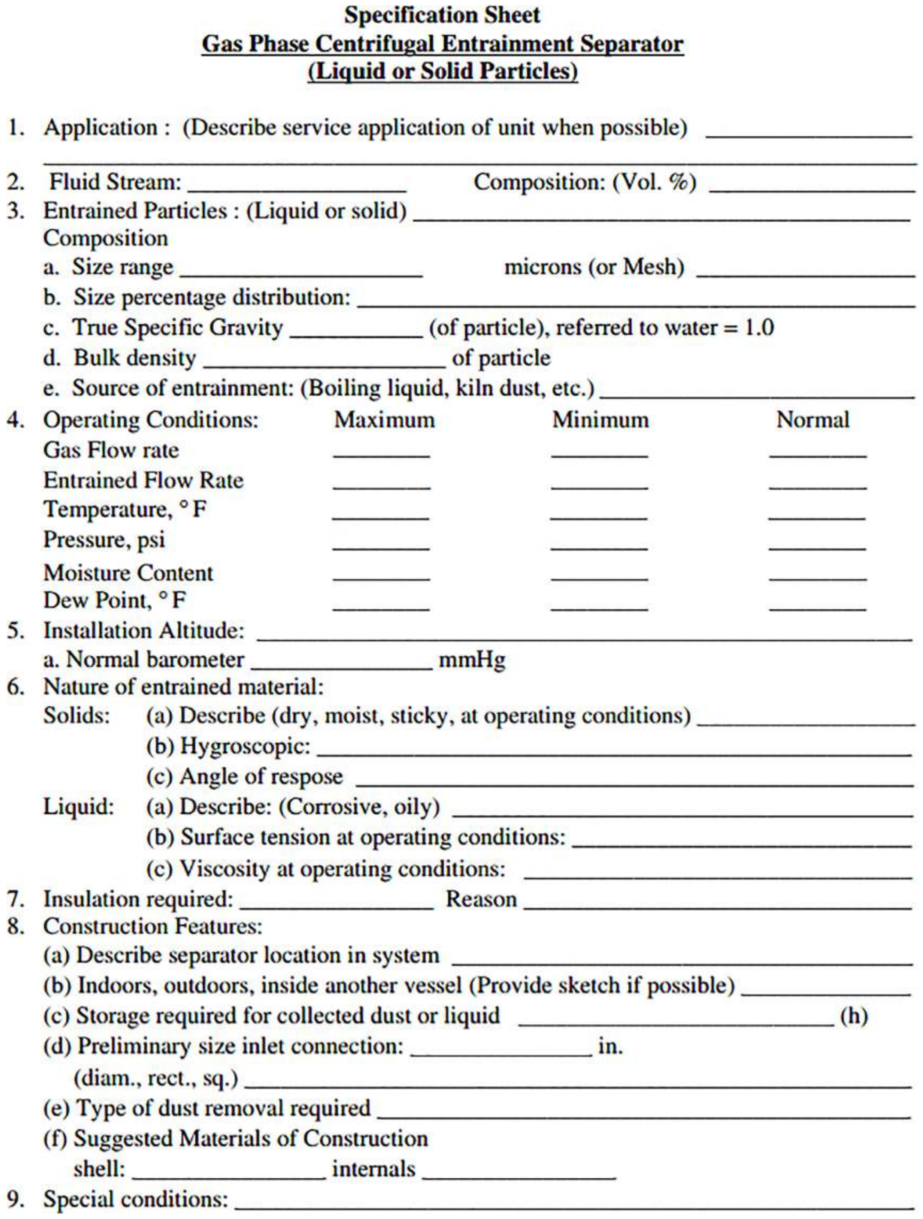
There are a few mechanical arrangements that use external power to exert centrifugal force on the gas particle stream. The fan type blades direct the separating particles to the collection outlet. Figures 18.107 and 18.108 show such a unit. These units are compact and have been used in various dust applications. However, caution should be used to avoid installations involving sticky or tacky materials which might adhere to the walls and blades of the unit. The efficiency of these is about 90–99% similar to a small, high pressure drop cyclone. The air handling performance can be predicted using the fan laws.

Scrubbers

Scrubber separators use a liquid to form some type of liquid surface (spray droplets, film, etc.) to assist the internal arrangements of the separator in the separating action. Essentially the incoming dust or liquid particles are wet by the action of the liquid (usually water or oil) and are made larger and/or heavier and thus can be separated from the moving stream. There are many types and styles of units falling in this classification (Figures 18.109–18.115a). Toy and Bonady [60] provide a good summary of manufacturers and their products for wet scrubbing.

One or more of the following mechanisms are employed in the separating action of the wet scrubbers.

1. Impingement—on internal parts.
2. Wetting—of particle to help agglomerate and prevent re-entrainment.
3. Diffusion—dust particles deposited on the liquid droplets. Predominant for the submicron and particles up to about 5 μm.
4. Humidification aids in flocculation and agglomeration of particles.
5. Condensation will cause particle size to grow if gas cooled below its dew-point.



Specification Sheet
Gas Phase Centrifugal Entrainment Separator
(Liquid or Solid Particles)

1. Application : (Describe service application of unit when possible) ________________
____
2. Fluid Stream: ________________ Composition: (Vol. %) ________________
3. Entrained Particles : (Liquid or solid) ________________________________
Composition
a. Size range ________________ microns (or Mesh) ________________
b. Size percentage distribution: ________________________________
c. True Specific Gravity __________ (of particle), referred to water = 1.0
d. Bulk density ________________ of particle
e. Source of entrainment: (Boiling liquid, kiln dust, etc.) ________________
4. Operating Conditions:

Operating Conditions:	Maximum	Minimum	Normal
Gas Flow rate	________	________	________
Entrained Flow Rate	________	________	________
Temperature, ° F	________	________	________
Pressure, psi	________	________	________
Moisture Content	________	________	________
Dew Point, ° F	________	________	________

5. Installation Altitude: ________________________________
a. Normal barometer ______________ mmHg
6. Nature of entrained material:
Solids: (a) Describe (dry, moist, sticky, at operating conditions) ________________
(b) Hygroscopic: ________________________________
(c) Angle of respose ________________________________
Liquid: (a) Describe: (Corrosive, oily) ________________________________
(b) Surface tension at operating conditions: ________________________
(c) Viscosity at operating conditions: ________________________
7. Insulation required: ______________ Reason ________________________
8. Construction Features:
(a) Describe separator location in system ________________________
(b) Indoors, outdoors, inside another vessel (Provide sketch if possible) ____________
(c) Storage required for collected dust or liquid ________________________ (h)
(d) Preliminary size inlet connection: ______________ in.
(diam., rect., sq.) ________________________________
(e) Type of dust removal required ________________________________
(f) Suggested Materials of Construction
shell: ______________ internals ______________
9. Special conditions: ________________________________

Figure 18.106 Specification Sheet, gas phase centrifugal entrainment separator (liquid or solid particles).

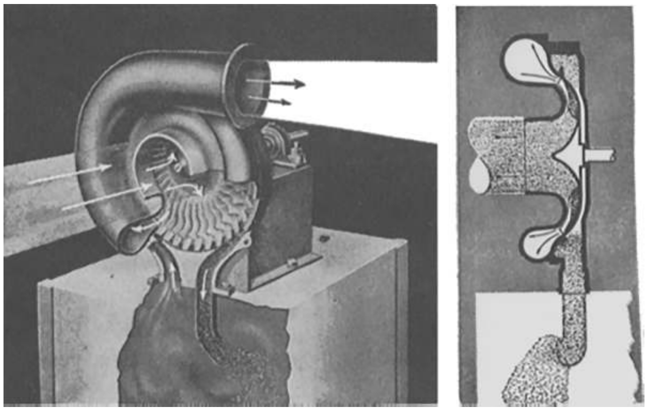


Figure 18.107 Inertial centrifugal dust separator (courtesy of American Air Filter Co.).

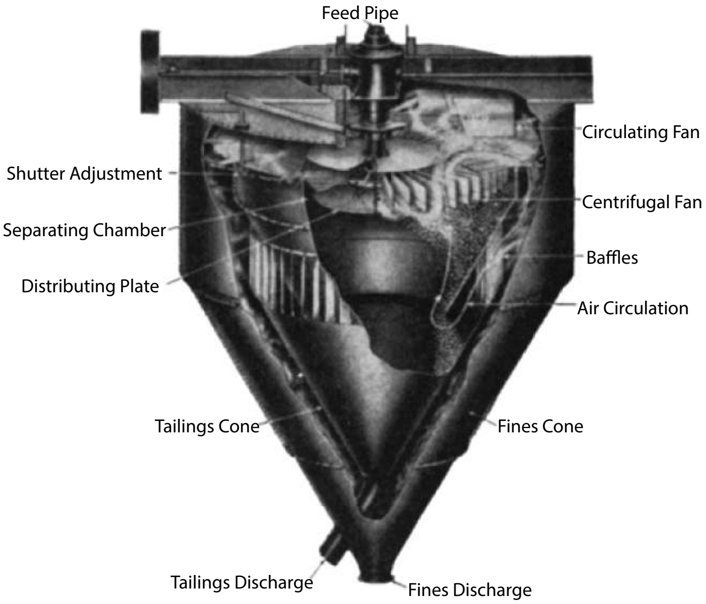


Figure 18.108 Inertial centrifugal dust separator (courtesy of Universal Road Machinery Co.).

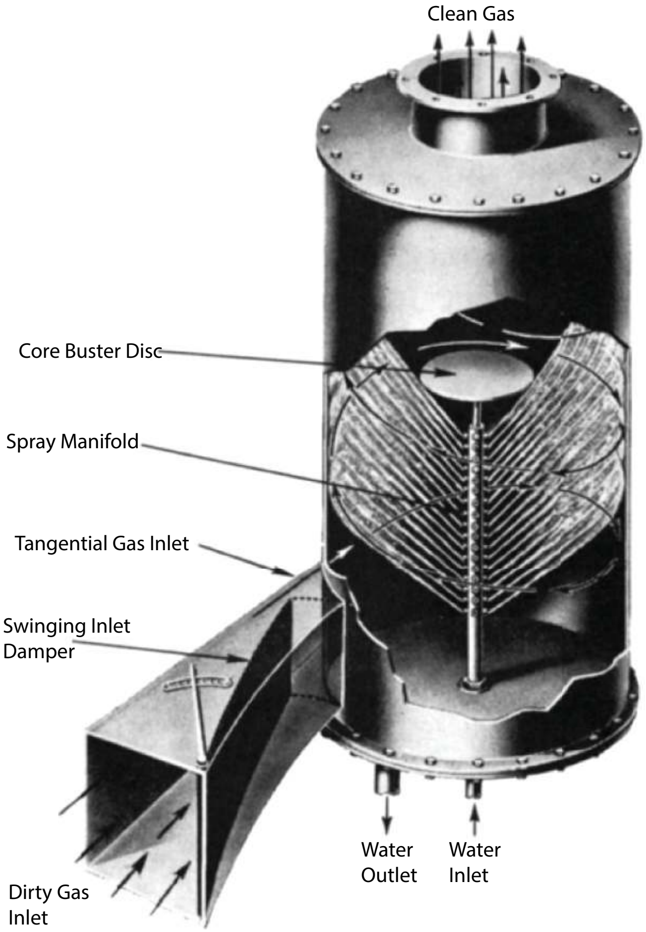


Figure 18.109 Cyclonic scrubber (courtesy of Chemical Construction Corp.).

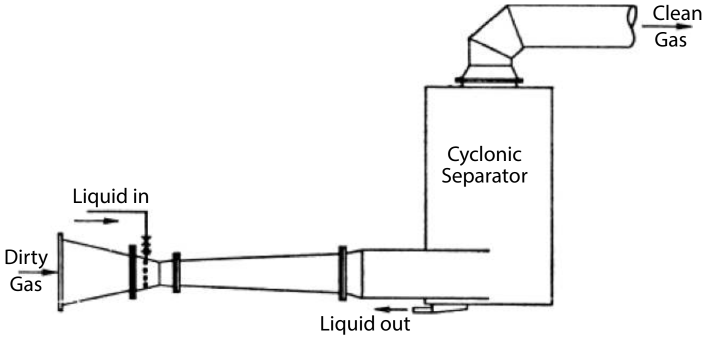


Figure 18.110 Venturi scrubber (courtesy of Chemical Construction Corp.).

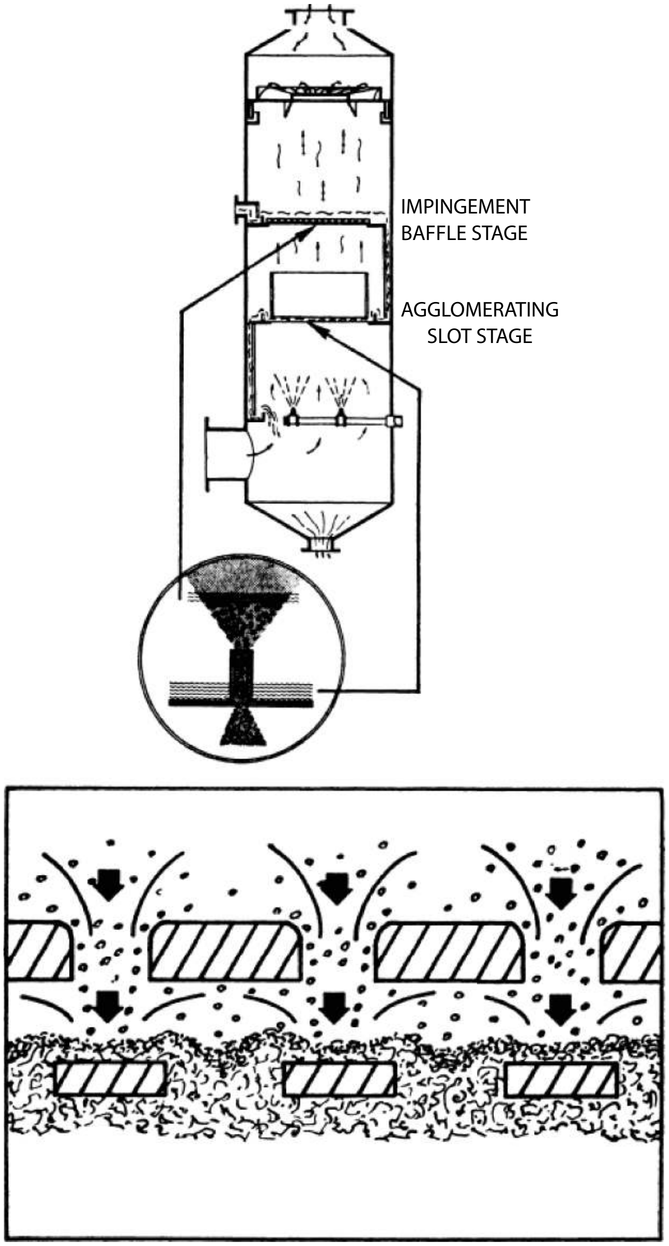


Figure 18.111 Impingement scrubber (courtesy of Peabody Engineering Corp.).

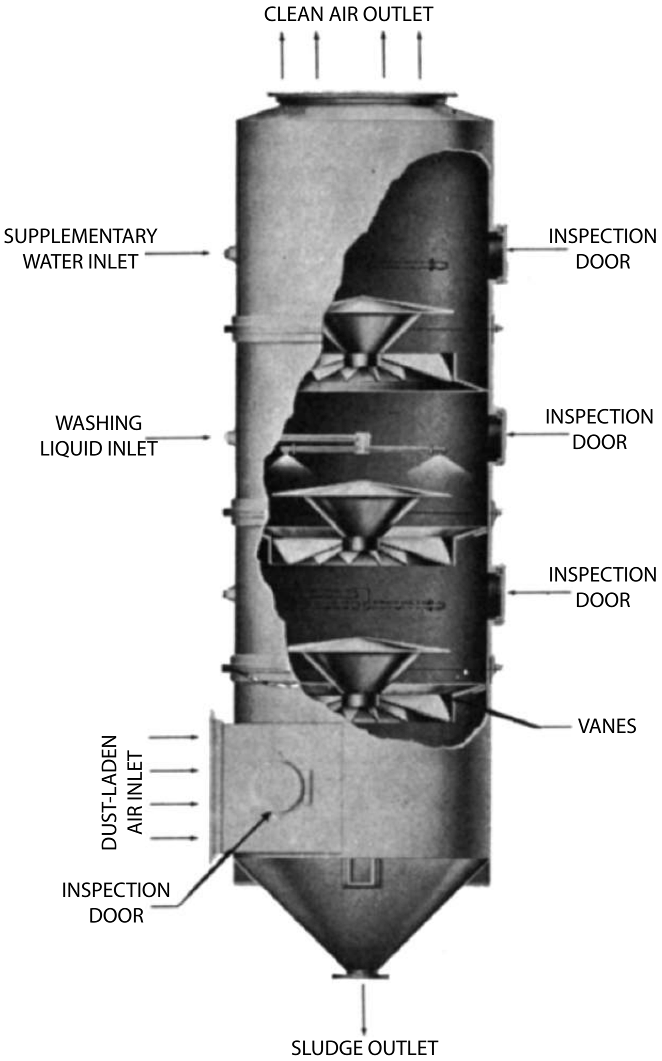


Figure 18.112 Spray scrubber (courtesy of the Ducon Co., Inc.).

6. Dust Disposal—running film action of liquid washes dust and collected liquid out of scrubber.
7. Gas Partition—segregates gas into small streams and segments when flowing through a liquid or foam.
8. Electrostatic Precipitation—the electrical charging of the liquid droplets may come about by the interaction of the gas and liquid streams. Not much known of this action.

The separating ability of most units is limited to 5 μm particles. However, some will take out 1 to 5 μm particles at a sacrifice in collection efficiency. Due to the peculiarities of each system as well as the equipment available to perform the separation, it is good to consult manufacturers regarding expected performance. Quite often they will want to run test units, particularly on difficult separations. Montrose [12] and Perry [5] give good descriptions of the various types of equipment illustrated in Figures 18.109–18.115a.

Figures 18.115a and b use a floating valve variable orifice opening as used in distillation contacting on the one or more trays included in the manufacturer's design. This provides for good contact to wet down the solid particles as well as scrub many water soluble gas/vapors in the incoming stream (such as chloride, sulfur, and nitrogen compounds). Heat and mass transfer can take place under these conditions. The pressure drop through this type unit typically ranges from 1–2 in. H_2O for a fivefold increase in gas flow rates. Particle removal can go as low as 0.5 μm to greater than 30 μm. Usually a wire mesh entrainment pad is mounted in the outgoing "clean" vapors to knock out liquid entrained particles, not solids.

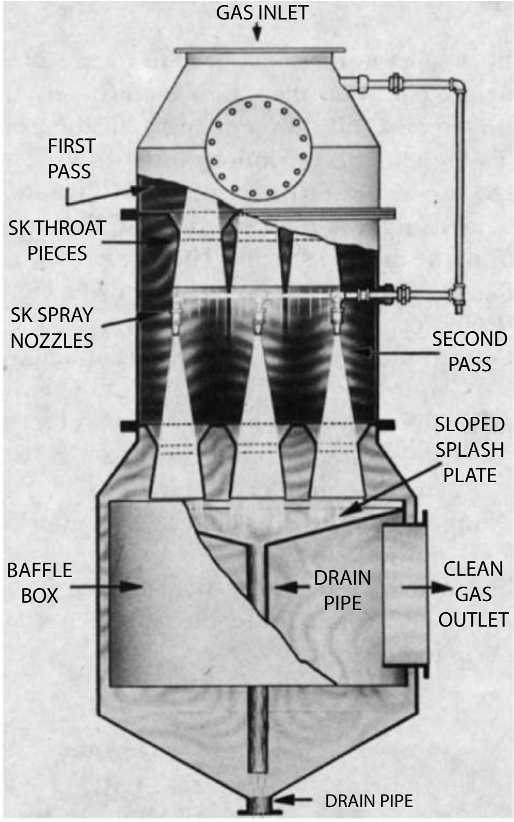


Figure 18.113 Spray scrubber–fume scrubber arranged for vertical down flow (courtesy of Schutte & Koerting Co.).

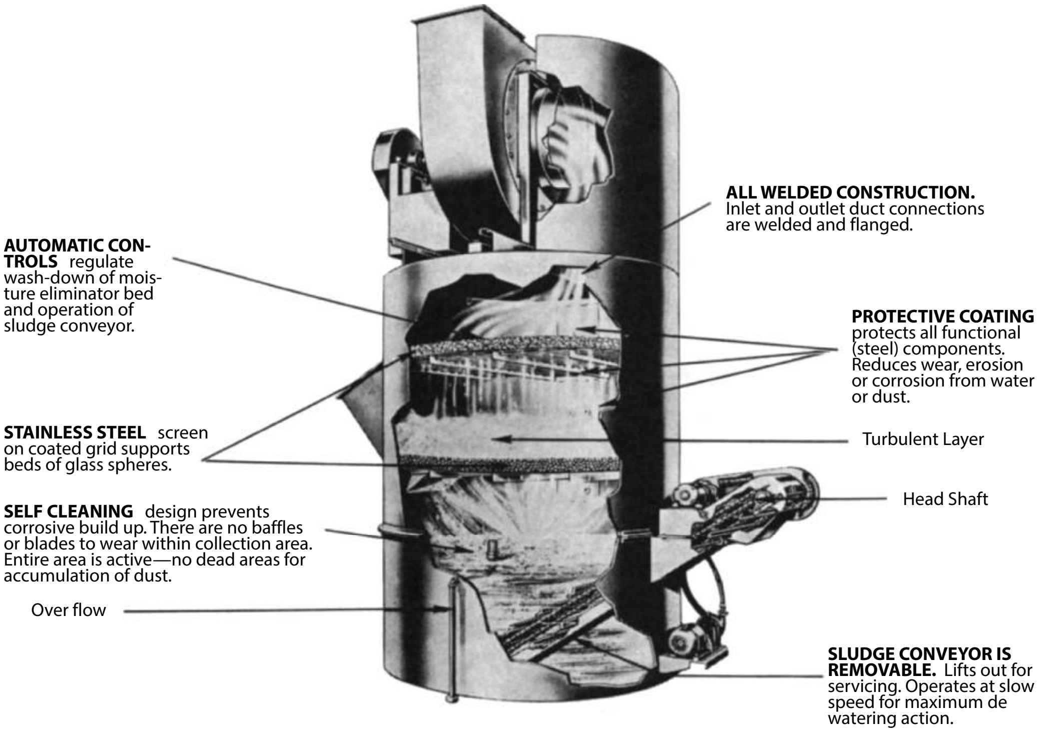
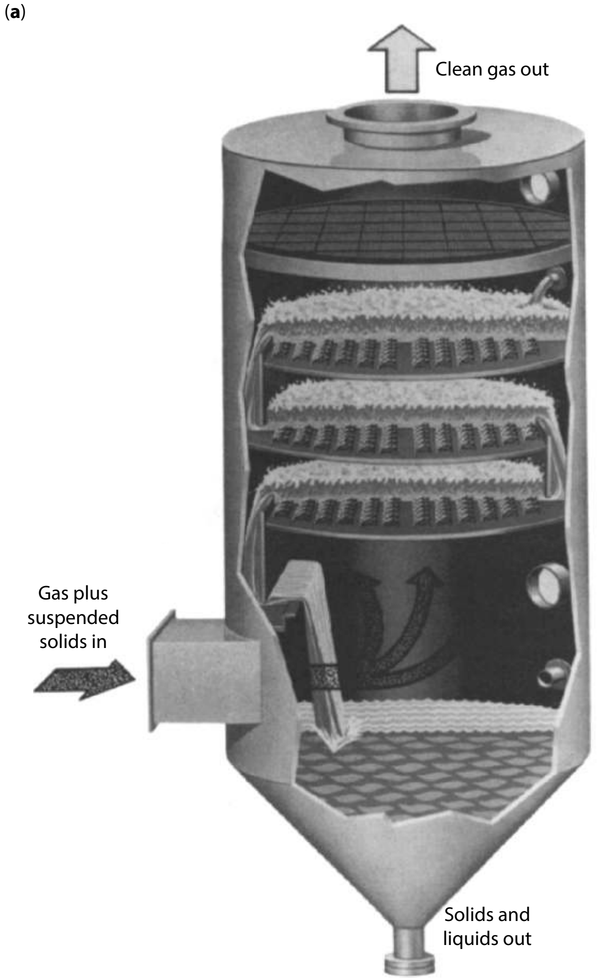


Figure 18.114 Tray-type scrubber with continuous sludge removal (courtesy of National Dust Collector Corp.).



The scrubber is comprised of one or more trays. Each tray contains numerous venturi openings. Each of the MultiVenturi openings is surmounted by a spider cage holding a floating Flexicap (see insert). In addition, each tray is equipped with one or more "downcomers" and weir flow baffles that control the scrubbing liquid as it flows across the tray and then to the tray below.

Figure 18.115a Variable orifice MultiVenturiFlexitray® scrubber at essentially constant pressure drop maintains good efficiency over wide flow rates (by permission from Koch Engineering Co. Inc.).

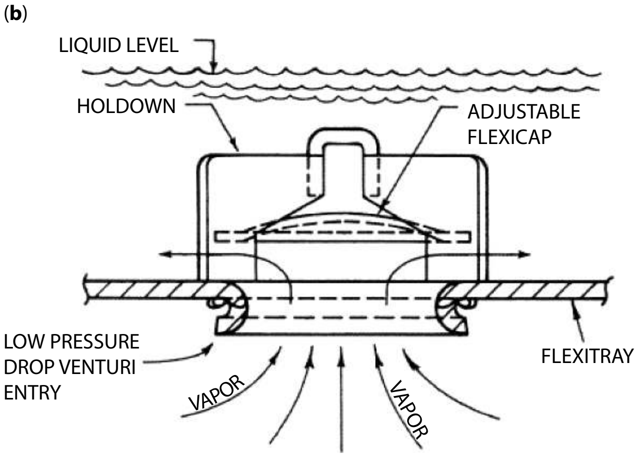


Figure 18.115b Adjustable "floating" caps for vapor flow (by permission from Koch Engineering Co. Inc.).

Cloth or Fabric Bag Separators or Filters

Beddow [61] provides additional details beyond the bag filter applications, and Boothroyd [62] provides a technical and analytical review of flowing gas–solid suspensions.

Filters of this type or class may be of the large bag filter type for large volumes of low plessure dust laden gases or vapor, or of the generally smaller cartridge or pack types for gas/vapors or liquids containing suspended solid materials.

Figures 18.116–18.118 show several units of the bag. The bags may be of cotton, wool, synthetic fiber, and glass or asbestos with temperature limits on such use as 180°F, 200°F, 275°F, 650°F, respectively, except for unusual materials (see Tables 18.12a and b). These units are used exclusively on dry solid particles in a gas stream, not being suitable for wet or moist applications. The gases pass through the woven filter cloth, depositing the dust on the surface. At intervals the unit is subject to a de-dusting action such as mechanical scraping, shaking or back-flow of clean air or gas to remove the dust from the cloth. The dust settles to the lower section of the unit and is removed. The separation efficiency may be 99%+, but is dependent upon the system and nature of the particles. For extremely fine particles a pre-coat of dry dust similar to that used in some wet filtrations may be required before re-establishing the process gas-dust flow.

For heavy dust loads these units are often preceded by a dry cyclone or other separator to reduce the total load on the bags. Suggested air-to-cloth ratios are given in Table 18.30.

Specifications

The details of specifications for bag filter dust collectors are important to a proper and operable design selection. There are many variables which must be furnished by the manufacturer so that the user can understand how the

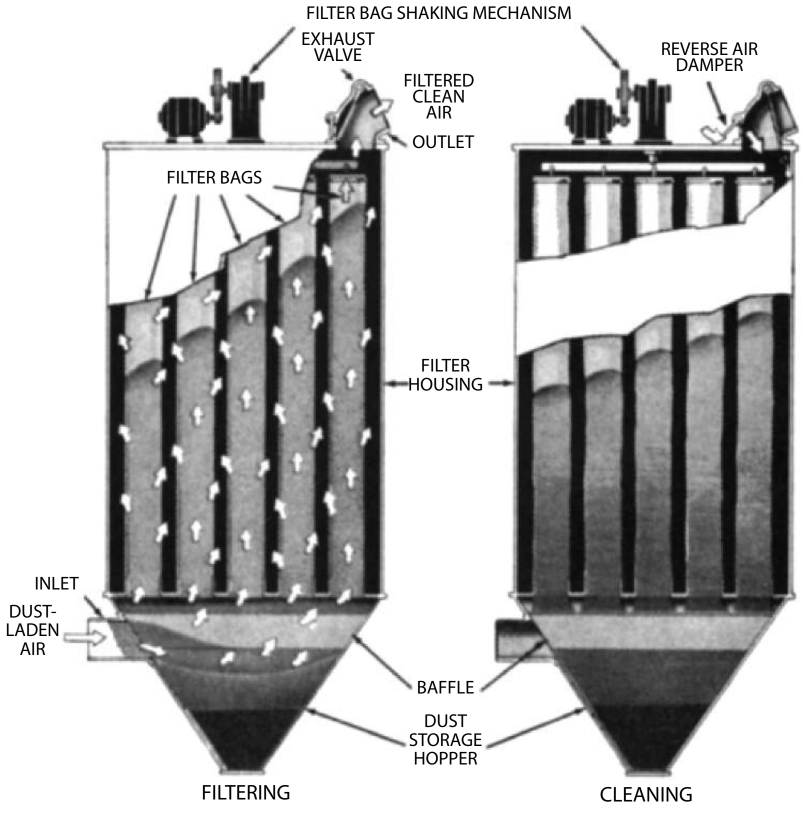


Figure 18.116 Bag filtration with mechanical shaking for bag cleaning (courtesy of Dracco Div. Fuller Co.).

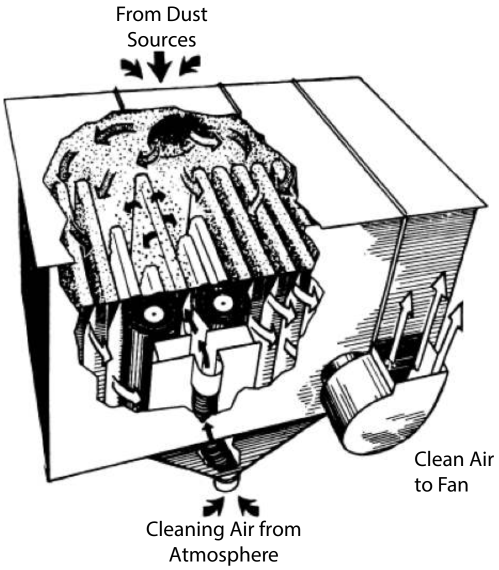
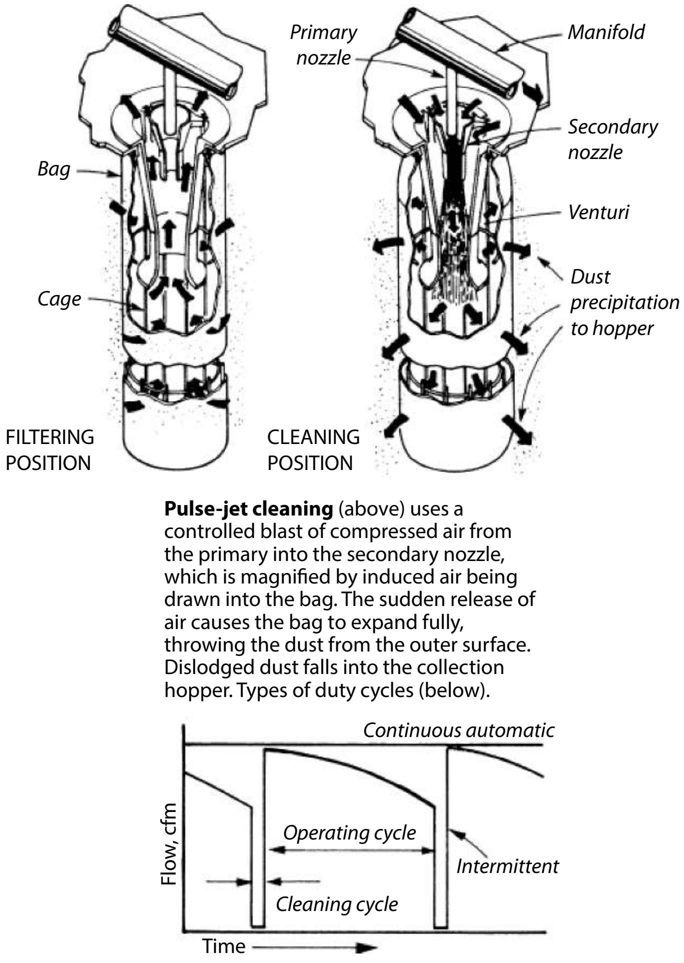


Figure 18.117 Bag filtration with continuous reverse air cleaning (courtesy of W. W. Sly Mfg. Co.).



Pulse-jet cleaning (above) uses a controlled blast of compressed air from the primary into the secondary nozzle, which is magnified by induced air being drawn into the bag. The sudden release of air causes the bag to expand fully, throwing the dust from the outer surface. Dislodged dust falls into the collection hopper. Types of duty cycles (below).

Figure 18.118 Pulse-jet air cleaning of fabric bags (by permission from Power, Nov. 1975, McGraw-Hill Co., Inc., New York, p. 41).

Table 18.30 Suggested air-to-cloth ratios for dust removal from air.

Type of dust	Ratio
Abrasives	2–2.5
Asbestos	2.5–3
Blast cleaning	3–3.5
Carbon	2–2.5
Cement (mills)	1.5–2
Cement (conveying and packing)	2–2.5
Clay	2–2.5
Coal	2–2.5
Feed	2.5–3
Graphite	1.5–2
Grinders	3–3.5
Gypsum	2–2.5
Lampblack	1.5–2
Limestone	2–2.5
Rubber	2–2.5
Salt	2.5–3
Silica Flour	2–2.5
Soap	2–2.5
Soapstone	2–2.5
Talc	2–2.5
Wood Flour	2–2.5

Note: Ratio is the volume in cubic feet per minute of dust-laden air to each square foot of active cloth area. If gain loading is above normal, ratios must be reduced (source: By permission, Bulletin 104. The W. W. Sly Mfg. Co., Cleveland, Ohio).

unit operates mechanically and the unit's dust loading capabilities. The larger the air/cloth ratio for the unit, the smaller will be its physical dimensions and generally, cost; however, the higher will be the frequency of cleaning. This can be quite troublesome; therefore low values of this ratio are preferable, consistent with the analysis of overall performance.

The removal or filtration of the entrained dust from the gas stream is accomplished by passing the mixture through a sufficiently porous fabric filter bag(s) (Table 18.31). These bags allow some air to flow through and are either cylindrical tubes or oblong tubes/bags. The dust is retained on the outside or inside (depending on unit design) of the bag surface and the small spaces between the fibers of the cloth (or felt). This dry cake builds up and acts as a pre-coat and then as the actual filtering medium as the dust particles build up. After a period of time, unique to the filter system of dust laden air plus bag type, the pressure drop will build up (These are low pressure and low pressure drop systems). Therefore, the dust or "cake" is removed (cleaned) from the outside of the bag by internal arrangements in the "bag house" or housing by such techniques as (1) shaking or vibrating the bag or bag assembly to drop the

Table 18.31 Physical properties of filter media.

	Cotton	**Wool**	**Glass**	**Polyester**	**Polypropylene**	**Nylon**	**Nomex (high temp. Nylon)**	**Rayon**	**Dynel**	**Teflon**	**Paper**	**Sintered metal**	**Woven wire cloth**	**Porous ceramics**
Particle Retention	2μ to >100μ	2μ to >100μ	3μ to >100μ	2μ to >100μ	2μ to >100μ	2μ to >100μ	2μ to >100μ	2μ to >100μ	2μ to >100μ	2μ to >100μ	3μ to >100μ	1μ to >100μ	2μ to >100μ	1μ to >100μ
Contaminant Holding Ability	G to E	E	F to G	G to E	F to E	F to E	F to E	F to E	F to E	G to E	G	F to G	F to G	F to G
Permeability	G	G	F	G	G	G	G	G	G	G	G	F	G	F
Chemical Compatibility	F	F	G	G	E	G	G	F	E	E	P	E	E	E
Temperature Limits (°F)	200	200	700	300	200	250	450	200	200	450	200	1200	1200	2000
Strength	G	F	G	G	G	G	G	F	G	G	P	E	E	G
Abrasion Resistance	G	G	P	E	E	E	E	G	G	E	F	E	E	G
Machineability (Workability)	E	E	F	E	E	E	E	G	E	E	F	F	F	F
Cleanability	G	G	P	G	E	G	G	G	E	E	P	P	G	P
Cost	L	L	M	M	L	M	M	L	L	H	L	H	M	H

Legend
P—poor F—fair G—good L—low M—medium H—high

dust into an integral hopper while there is no-flow of air-dust feed into the unit or compartment, or (2) back pulse with jets of air into each bag (Figure 18.118). The criterion should be a constant pressure drop across the fabric for a fixed air flow and a specified dust loading [63]. The air-to-cloth ratio so often used is only useful when comparing a particular manufacturer's equipment for handling different materials, and not for comparing manufacturers. Kraus [64] provides an excellent summary of many details associated with specifying and selecting bag filters.

The following are suggested filter specifications:

1. Performance: define air/gas and dust rates, particle size distribution, and percent of particle sizes.
   a. Temperature at inlet to baghouse
   b. Moisture concentration, dewpoint
   c. Chemical composition of vapor and of dust, including any abrasive, hydroscopic or other characteristics.
2. Define dust recovery, as percentage below a certain particle size.
3. Indicate, if known, preferred bad material that will withstand environment, e.g., fibers of glass, polyester, Teflon®, Nomex®, polypropylene, polyethylene, cotton, wool, nylon, Orlon®, Dacron®, and Dynel®. The type of weave of fibers should be recommended by the manufacturer. The fabrics may be felted or woven [63, 65] in weaves of plain, satin, or twill, and should be resistant to any corrosive material in the solid particles or the gas stream.
4. Manufacturer should recommend
   a. bag size (diameter, length)
   b. bag holding hardware: anti-collapse spreader rings, snap rings, and so on.
   c. number of baghouse compartments, number of bags per compartment.
   d. air/gas flow cycle to compartments.
   e. complete description with mechanical details of bag cleaning system (shaking, air-jet, and so on.)
   f. dust removal arrangement.

The cleaning system set up for a particular bag house will determine whether the filtering system operates continuously or batch/intermittently. Some systems operate as a continuous batch, with sections of the entering chambers being isolated by valving to automatically switch from one section of one bag to another. Thereby, one or more bag groups/sections filter while another one or more are not operating, but are in the dust removal cycle. The permeability of the fabric is generally stated as the clean airflow in (cubic feet/min/square foot of fabric at a pressure differential of 0.5 in. H_2O as determined by the ASTM standard D-737 (Frazier Test) [63]. Whereas this test is useful, several fabrics may have the same permeability yet have different fiber surfaces, and thereby do not perform the same for a specific application.

The felted fabrics are generally used for maximum recovery of product and are used at high face velocity for air-flow-to-cloth-area ratio. The felt promotes the greatest dust collection surface. Monofilament fibers require special attention to ensure a uniform open space between the filaments.

The woven fabrics have various yarn patterns for different spacings between the yarn fabrics (Table 18.31). There is a wide variety of choices for not only the materials of construction but the tightness of weave and the size of the yarn. All of these factors along with the others noted earlier, make the selection of bag fabric an art that requires manufacturer's and plant's actual field tests. Woven fabrics have a low ratio of weave openings for yarn area and generally have a limited face velocity for air flow of about 1.5 to 3.0 $ft^3/min/ft^2$ [63].

Newer fabrics, not in common use but in development, test, and field trials, are described for higher temperature applications by Parkinson [66]. Application to 400°–2100°F are potentially available using ceramic fibers Nextel 312®, laminated membrane of expanded PTFE on a substrate, polyimide fiber P-84, Ryton® polyphenylene sulfide, and woven fiberglass. The heat and acid resistance of these new materials varies, however they can be handled at higher temperatures.

New cartridge designs for bag houses will allow improved servicing and cleaning techniques. It is important to keep bolts, nuts and other potentially loose items to a minimum inside the unit, as vibration from air/gas flow and bag cleaning can loosen nuts, break small welds, and ultimately tear holes or rip bags. The bag construction is likewise extremely important, since loose edges and "unlocked" seams will fray and tear, allowing fibers into the product dust. The bag construction must have straight seams in order for them to bend and flex properly on cleaning and/or loading.

Cartridge filters may be single units or clusters in a single container or canister. Figures 18.119–18.125 illustrate typical units. These may be designed to filter suspended solids from gases or liquids. Table 18.31 presents representative physical property and application data for the more commonly used filter media. These media may be in filament, fiber, or "felt" form and arranged by weaving techniques to control the pore or free spaces to specific size for removal

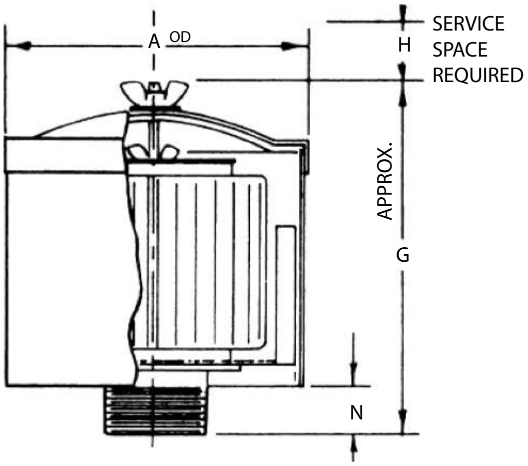


Figure 18.119 Typical blower intake filter-silencer. Air to blower leaves through pipe connection, which may be screwed or flanged (courtesy of Dollinger Corp.).

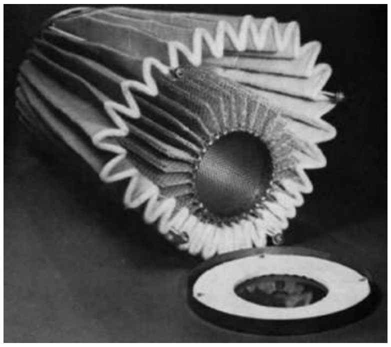


Figure 18.120 Pleated radial-fan filter cartridge. Filtration is from outside to inside (courtesy of Dollinger Corp.).

Figure 18.121 Wound filter tube on stainless steel core (courtesy of Filterite Corp.).

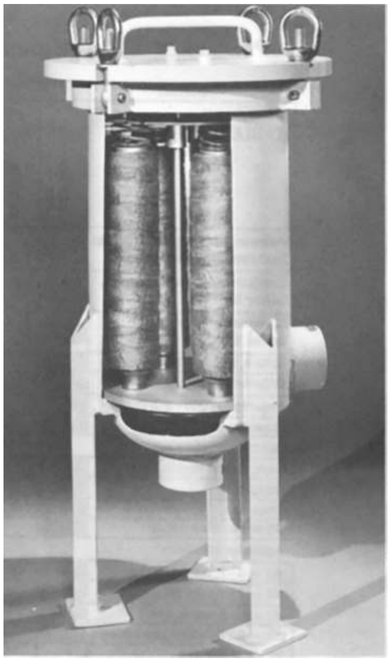


Figure 18.122 Cluster of filter cartridges in a single chamber (courtesy of Filterite Corp.).

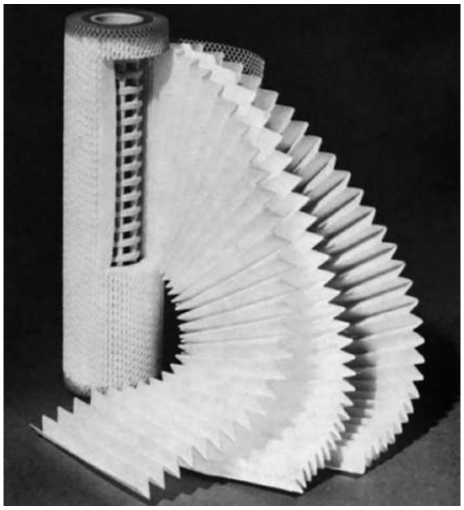
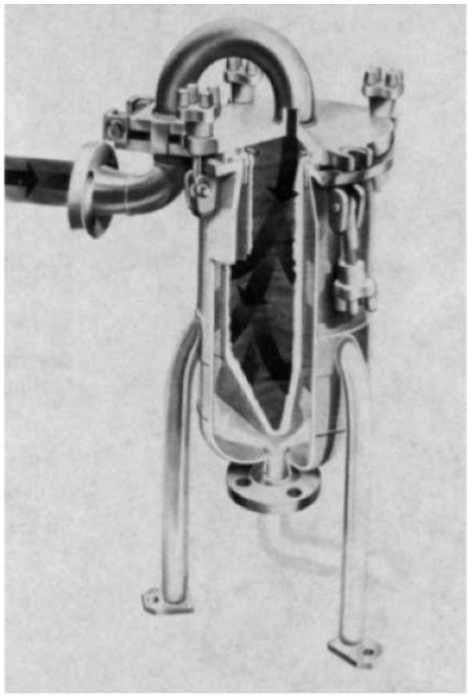


Figure 18.123 Cartridge-type filter-pleated membrane (courtesy of Gelman Instrument Corp.).

of various sizes of particles. The particle size retention listed in the table ranges from 0.006 μm to over 100 μm. A micron is often termed "micrometer" or a millionth of a meter, using symbol μm.

Cartridges as illustrated are considered "throw-away" and are removed from service when the pressure drop builds up to a predetermined value, or when the effluent changes color or becomes opaque with suspended material breaking through. The flow in most applications is from outside cartridge to inside and into the hollow metal or plastic collection tubes. It then flows into the outlet pipe to the process. Materials for these cartridges are most commonly selected from cellulose, glass fibers, polypropylene (woven and non-woven) fibers, or monofilaments, molded resins, ceramics, or resin-impregnated fiberglass. The last three are termed "depth" filters, as they can hold a large amount of solids before the pressure drop builds up excessively. "Surface" filters are usually made of paper, non-woven fabrics, or cast membranes and are usually pleated to provide more working surface area. This type is fabricated from sheets of porous non-woven fabric often used for the absolute capture of sub-micron particles and has a sharp cutoff in particle size retention [67]. Yarn wound filters often have a graded density or decreasing pore size structure.



This three-dimensional cutaway drawing illustrates the filtering operation of the GAF® filter-bag pressure filter system, showing the flow patterns of unfiltered liquid through a preselected micronrated felt filter bag which renders the desired quality of filtered product.

Figure 18.124 Flow scheme for GAF filter-bag pressure filter system for liquids (courtesy of GAF Corporation, Chemical Group, Greenwich. Co.).

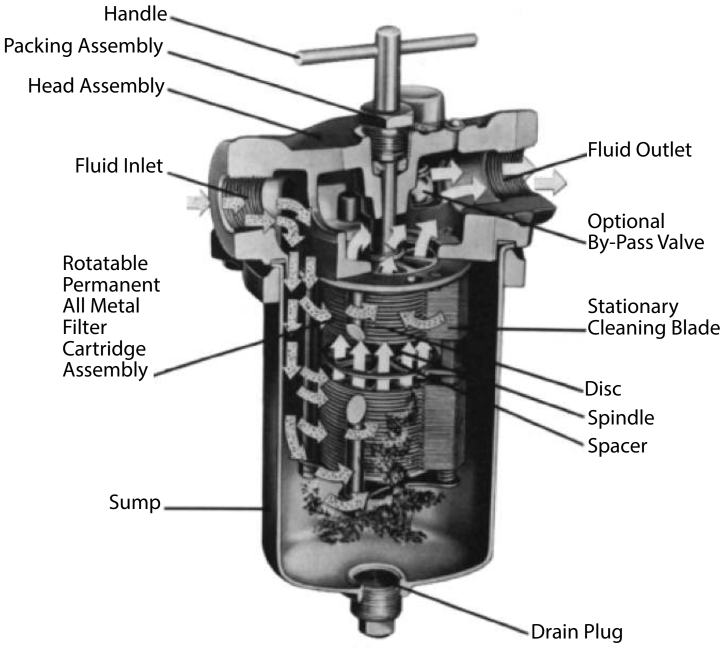


Figure 18.125 Edge-type filter with the external online cleaning (courtesy of AMF Corp., Cuno Div.).

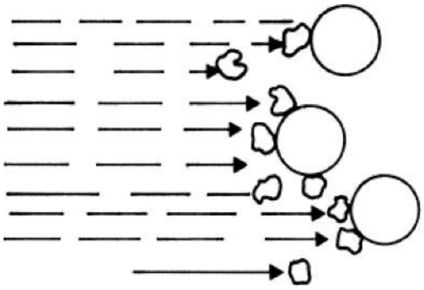
To aid in selection of the most probable successful filter media for the service, the summaries of Tables 18.12a and b can be a useful guide [68]; however, for some applications, actual testing in the plant using plant fluid streams can be the most conclusive. This plant testing is not necessary for every situation because the manufacturer has large data files to often aid in a good selection. Generally the ability to collect solids at low flow rates is greater for the wound filter.

Because the suspended particles are "captured" by different physical mechanisms depending on the particle size, shape, density, and concentration, all cartridges do not perform the same. The "capture" may be by (1) direct interception, (2) sieving, and/or (3) bridging [69] (see Figure 18.126). The cartridges from one manufacturer are generally consistent in performance; however, all cartridges from just any manufacturer may not be interchangeable in performance.

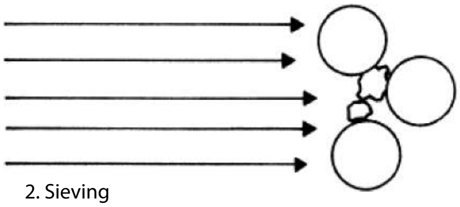
The micron ratings of a cartridge are intended to indicate the smallest particle that will be retained by the pores of the filter element. Often a "rough-cut" pre-filter is installed ahead of a final or "polishing" filter in order to increase the life of the final unit. Unfortunately, the method for determining the micron rating is not a universal standard between manufacturers. Thus, one manufacturer's "50 μm" filter may not perform the same as another manufacturer's with the same rating number. The only reliable approach is to send the manufacturer an actual sample of the fluid and let it be tested in order to select the filter for the job. Alternatively test the unit in your plant's field application [68].

An important feature of these cartridge units is the mechanism for assembling one or more in the housing. The top/bottom sealing mechanisms determine what style of cartridge is required (open both ends, open one end) and the method of pressure loading/sealing each cartridge into its bracket in the housing. The housing may hold one or 40 cartridges, and the assembly inside to prevent leakage and cross-contamination is essential to good performance as a filtering device. The housings can be made of various metals (carbon steel, stainless, alloy) or plastic-lined steel using corrosion resistant polymers, or elastomers, or solid plastic.

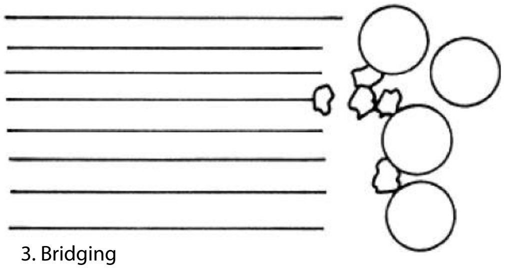
The cartridges can be selected to be useful over the range of low to high viscosities, that is, 100,000 cP with temperature ranges to 750°F at higher pressure of up to 3,000 psi [68]. Usually for the average application, the concentration



1. Direct Interception



2. Sieving



3. Bridging

Figure 18.126 "Capture" mechanism for cartridge filters (adapted by permission from Shucosky, A. C., Chem. Eng., Vol. 95, No. 1, 1988, p 72).

of the suspended solids is not over 100 ppm, but can be higher. These units do not perform well with pressure pulsations or surges in the system. Note the differences in expected performance of Figure 18.127 between a pleated cartridge. This does not necessarily mean that all cartridges perform in this manner, but these are typical of expected performance curves. When examining particle retention ratings, examine Shucosky [69].

Note: (a) Designations for both nominal and absolute ratings are based on the measure of a particle size, not a pore size. (b) Ratings are based on arbitrary laboratory tests by the filter manufacturer and can vary in actual plant conditions as previously discussed.

For some critical applications (such as polymer melt, beverage, or pharmaceutical filtration), it may be important to avoid cartridges that have a “nap” or “fuzz” on the fiber used, because these extremely fine fibers tend to break off and drift through the cartridge and go out with the finished product, thereby creating a visual acceptance problem, if not outright contamination.

In actual practice some companies have cartridges that will remove to 0.25 μm in size. The smaller the particle size that is specified to be removed from the vapor or liquid, the higher will normally be the ultimate pressure drop or lower the holding capacity. In normal operation, the pressure drop initially is quite low, perhaps 1 to 3 psig depending on flow rate, but as the solids build up, the pressure drop will rise to 10 to 35 psig, in which range most companies recommend replacement.

These replaceable cartridges or packs are the most commonly used; however, there are cartridges of wire mesh, sintered or porous metal which can be removed, cleaned, and replaced. Usually, the fine pores of the metal become progressively plugged and the cartridges lose capacity. They are often used for filtering hot fluids, or polymers with suspended particles, pharmaceuticals, and foods (liquids). In the case of polymers and other applications a special solvent and blow-back cleaning system may be employed.

The small cartridge units can be conveniently placed ahead of instruments, close-clearanced pumps, or a process to remove last indications of impurities in suspension.

Other useful cartridges are as follows:

1. woven stainless steel (or other wire) wire screen mesh (Figures 18.128 and 18.129).
2. wire wound (Figure 18.128).
3. sintered metal (Figure 18.130).

The woven wire mesh types are formed to control the open space between the wires, thereby limiting the maximum size particle that can pass through. The cartridge is installed in cases or small vessels to facilitate quick replacement,

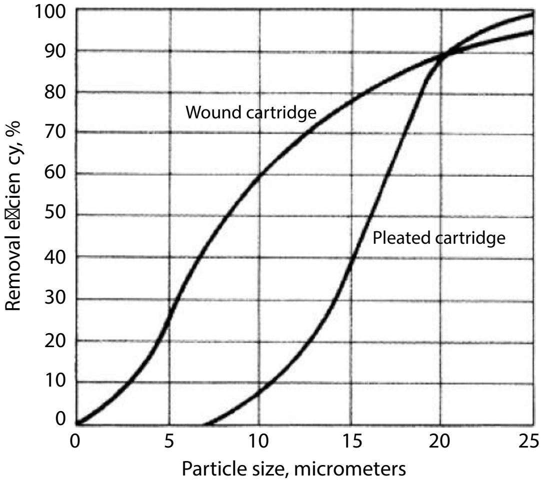


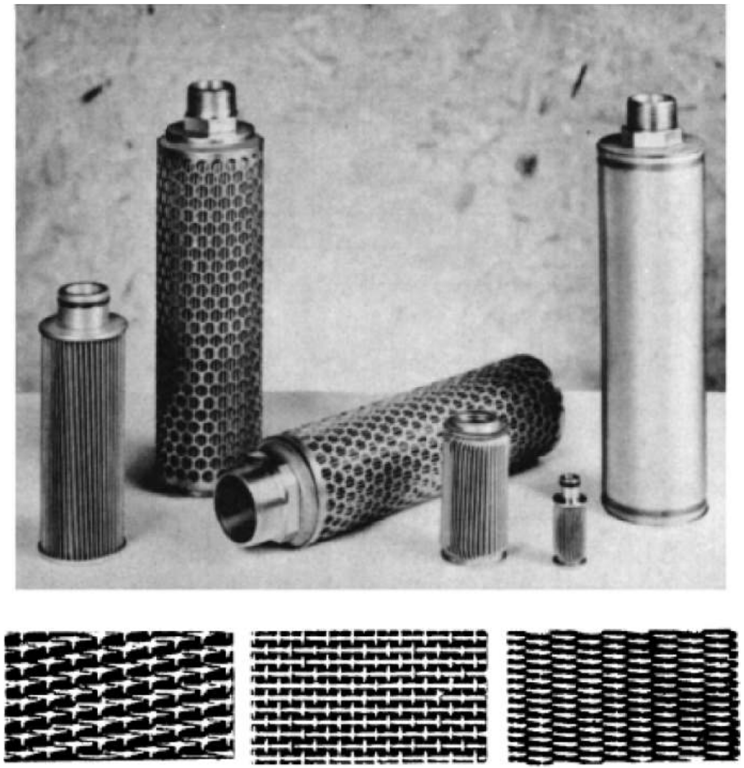
Figure 18.127 Pleated and wound cartridges differ in removal-efficiency profile (by permission from Shucosky, A. C., Chem. Eng., Vol. 95, No. 1, 1988, p. 72).

or they can be arranged for backwash by use of proper piping connections. The wire-wound units have consistent spaces for uniform particle size filtering.

The sintered metal units have uniform permeability with void spaces approximately 50% by volume for some metals and manufacturing techniques. The pore sizes can be graded to remove particles from 1 μm to 20 μm for liquids and smaller sizes when used in gaseous systems (see Figures 18.130). Metals usually used are stainless steel, nickel, monel, inconel, high nickel alloys, and special designs for unique services.

The pressure drops for these units are typically low, ranging from 0.2 to 10–15 psi. The woven wire mesh runs even lower in pressure drops for the same or larger flow rates. Consult the manufacturers for specific application data.

With some types of particles the porous metal tends to plug, but they can usually be backwashed or washed with a solvent or acid/alkali to remove the particles from within the metal pores. This is one reason why manufacturer's testing or plant testing can be important to the proper selection. Once the internal plug-gage has reached a point of reduction in flow-through capacity, it must be discarded. The actual cost of this type of cartridge is several times that of the non-metallic unit; therefore, the economics involving the life span of each unit should be examined.



DUTCH TWILL PLAIN OR SQUARE DUTCH WEAVE

Nominal Rating Microns	Equivalent Standard Absolute Rating Microns
2	20
5	25
10	40
20	55
40	90

Figure 18.128 Woven wire mesh filter cartridges (by permission from AMF Corp., Cuno Div., Catalog MP-20.1).

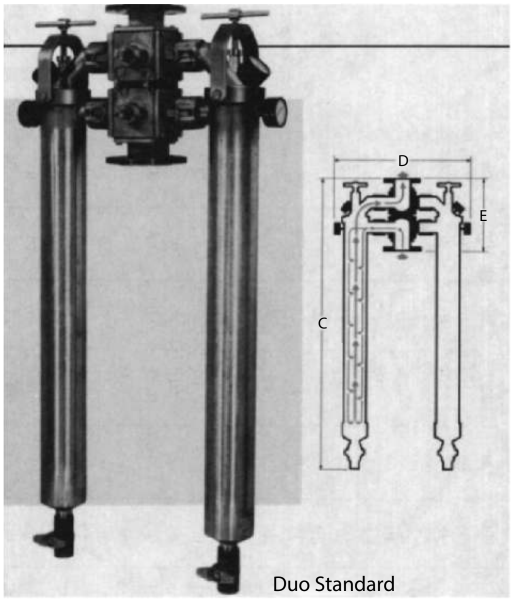


Figure 18.129 Tubular in-line pressure filter with reusable elements. The flow: unfiltered liquid enters the inlet port, flows upward, around, and through the media, which is a stainless steel or fabric screen reinforced by a perforated stainless-steel backing. Filtered liquid discharges through the outlet (top) port. Because of outside-to-inside flow path, solids collect on the outside of the element so screens are easy to clean (by permission from Ronningten-Petter® Engineered Filter Systems, Bulletin RP-2).

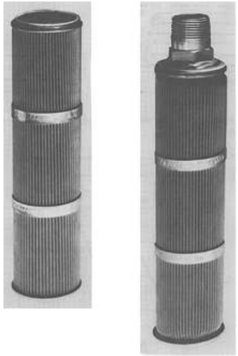


Figure 18.130 Porous sintered metal filter elements (by permission from Pall Process Filtration Co.).

Electrical/Electrostatic Precipitators

The electrical precipitator is a dry dust or liquid mist removal unit which utilizes the ionization of the process gas (usually air) to impart electrical charges on the suspended entrained particles and effect particle collection by attraction to an oppositely charged plate or pipe. This type of unit is in use in services which are difficult for other types of entrainment removal equipment. Figures 18.131–18.133 illustrate the usual fundamental action of these units.

For these units the usual particle size for removal is greater than 2 μm with a loading rate of greater than 0.1 grains/ft^3, with a collection efficiency of 99%±. The pressure drop is very low for a range of gas velocity through the unit of 100–600 ft/min (31–183 m/min) [70]. Operating temperatures can be as high as 1000°F (538°C) and above [71].

To improve the efficiency of collection, several units can be installed in series. The plate type unit is the most common design for dry dust removal, while pipe design is mainly for removal of liquid or sludge particles and volatilized fumes. The plates/pipes are the collecting electrodes, with the discharge electrodes suspended between the plates or suspended in the pipes [72–74].

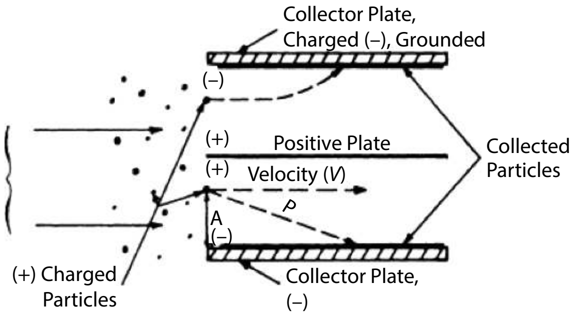


Figure 18.131 Charging particles in electrostatic precipitator (by permission from, adapted after, A. Nutting, American Air Filter Co.).

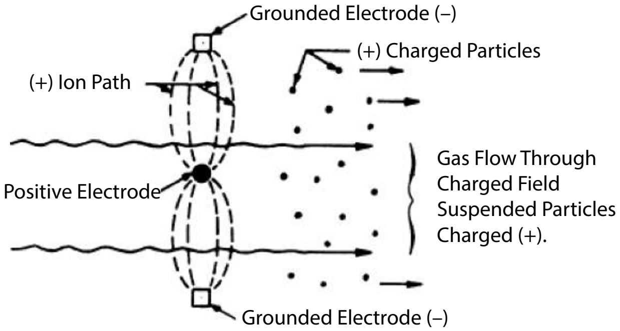


Figure 18.132 Particle collection (by permission from Nutting, A., American Air Filter Co.).

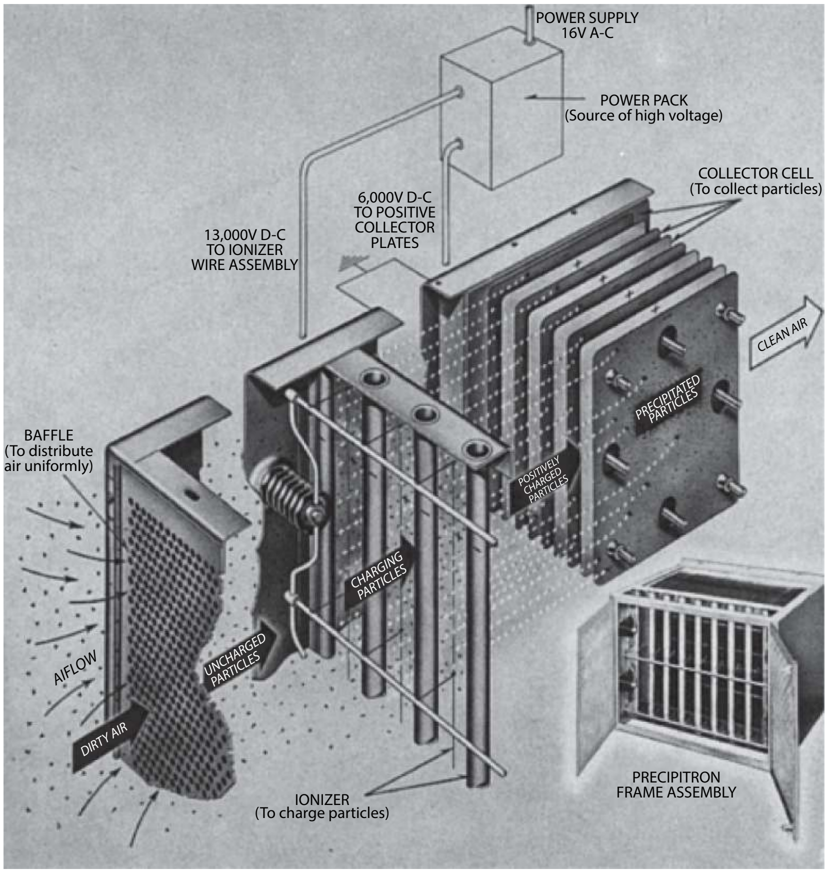


Figure 18.133 Electrical precipitator principle of operation (courtesy of Sturtevant Div. Westinghouse Electric Corp.).

In operation, the voltage difference between the discharge and collecting electrodes sets up a strong electrical field between them [74]. The "dirty" gas with particles passes through this field, and the gas ions from the discharge electrode attach to the suspended "dirty" particles, giving them a negative charge. The charged particles are then attracted to the positively charged collecting electrode, discharging their charge on contact, becoming electrically inert.

Collected liquids flow down the pipes and drain to a collection sump. Collected solids are washed off the plates with water or other liquid. Sometimes the dust/solids can be removed by mechanically vibrating or knocking on the plates while the particles are dry. The electrical power of the precipitator is applied only to the particles collected, thereby allowing for large volumes of gas to be handled with very low pressure drop.

For corrosive gases/liquid particles, corrosion resistant metals can be used for construction.

The performance of the unit involves the gas characteristics, analysis, velocity, flow rate, dust or liquid particle size and analysis, resistivity, and required final particle efficiency of removal. Some particle materials of high electrical resistivity prevent proper electrical operation.

Table 18.32 illustrates some industrial application of electrostatic precipitators; however, it is not intended to be all inclusive.

Electrostatic Precipitator Explosion: A Case Study of an Explosion in the ExxonMobil Torrance, California Refinery's Electrostatic Precipitator (ESP) Control Air Pollution due to a Lacked Safety Instrumentation, Equipment Failure, Safe Operating Limits and Improper Safeguard as Sufficient Hazard Analysis

On February 18, 2015, an explosion at the ExxonMobil refinery in Torrance, California. It occurred in the refinery's fluid catalytic cracking (FCC) unit, where a variety of products, mainly gasoline are produced. A reaction between hydrocarbons and catalyst took place in the hydrocarbon side of the FCC unit. The remainder of the unit comprises of a portion of the reaction process and a series of pollution control equipment that uses air referred to as the "air side" of the unit.

It is critical that hydrocarbons do not flow into the air side of the FCC unit as this can create an explosive atmosphere. On the day of the incident, a slide valve that acted as a barrier failed that allowed hydrocarbons to flow into the air side of the FCC where they ignited in a piece of equipment called the electrostatic precipitator (ESP) control air pollution causing an explosion of the ESP. Figure 18.134 shows the piping connection of the FCC, the main distillation units and the ESP; Figure 18.135 illustrates these main units.

Process Description

The FCC unit cracks heavy, high boiling point hydrocarbon molecules into smaller molecules with lower boiling points. The main product produced by the FCC unit is gasoline. Figure 18.136 shows a schematic of Torrance FCC unit. A large portion of the FCC unit was involved in the sequence of events that resulted in the explosion in the unit's ESP control air pollution, which removed catalyst particles from the regenerator combustion gas to meet environmental regulations before it was discharged into the atmosphere. During normal operation, a catalyst in the form of small spherical particles circulates between the reactor and the regenerator in the direction of the circular arrow as shown in Figure 18.137. The catalyst is typically fluidized implying that the solid catalyst is aerated with hydrocarbon vapor, steam or air so that it behaves like a liquid (Figure 18.138). The catalyst both drives the cracking reaction and transfers heat from the regenerator to the heavy hydrocarbon feed entering the reactor riser. Figure 18.139 shows a photograph of the catalyst used in the FCC unit.

The cracking reaction occurs in the reactor riser (Figure 18.137b). Heavy hydrocarbons are fed into the reactor riser, where they vaporize upon contact with fluidized hot catalyst and the cracking reaction begins (see Chapter 8, Vol. 1 of the series). The cracking reaction continues as the mixture of hydrocarbon vapor and catalyst travel up the riser. Coke, a solid byproduct of the cracking reaction, deposits onto the catalyst particles during the reaction process, making the catalyst less effective, thereby reducing the activity of catalyst. The coke-covered catalyst is referred to as

Table 18.32 Precipitator operating data for common CPI applications.

Precipitator operating data for common CPI applications	Ventilating				Grinding		Drying						
	Asphalt converter saturator	**Glass melting**	**Pot line**	**Carbon plant**	**Cement finish grind**	**Jet pulverized catalyst**	**Cement dryers**	**Bauxite dryers**	**Gypsum dryers**	**Copper dryers**	**Other drying uses**	**Dry process cement**
Gas flow, 1,000 cfm.	5.0 to 24.8	7.0 to 10.0	94 to 120	15.5 to 27.0	10.0 to 13.5	10.0 to 13.5	40 to 80	100 to 150	11 to 150	39 to 45	20 to 187	62 to 305
Gas temperature, F	100 to 150	525 to 550	190 to 220	53 to 80	75 to 200	75 to 200	100 to 280	150 to 250	130 to 200	350 to 400	105 to 400	600 to 750
Gas pressure, in. water	−4 to −2	−6 to +6	Negative	Negative	−10 to +10	−10 to +10	−10 to +10	−10 to +10	Negative	–	−10 to +10	−4 to +4
Gas moisture, volume %	5	Ambient	2 to 3	Ambient	1 to 10	1 to 15	5 to 12	30 to 35	5 to 15	10 to 20	15 to 100	5.7 to 15.0
Inlet dust concentration, grs/cu ft	0.1 to 0.5	0.66 to 0.80	0.07 to 0.16	7.0 to 17.2	29 to 70	29 to 70	20 to 40	20 to 40	6.5 to 150	50 to 52	0.05 to 40	3.7 to 43.0
Outlet dust concentration, grs/cu ft	0.02 to 0.05	0.13 to 0.03	0.003 to 0.007	0.03 to 0.13	0.021 to 0.044	0.021 to 0.044	0.05 to 0.5	0.02 to 0.07	0.10 to 0.30	0.025 to 0.04	0.02 to 0.4	0.001 to 0.938
Power input, kw	3 to 7	7 to 12	10 to 14	3 to 4	4 to 6	4 to 6	4 to 12	15 to 20	7 to 12	21	7 to 24	2.4 to 51.0
Collection efficiency, %	90 to 98	98 to 98.5	86 to 97.6	98.5 to 99.8	95 to 99.96	95 to 99.96	90 to 99	99+	97 to 99.7	99.0 to 99.5	90 to 99	80 to 99.97

(*Continued*)

Table 18.32 Precipitator operating data for common CPI applications. (*Continued*)

Calcining					**Metal acid recovery**			**Miscellaneous**							
Wet process cement	**Gypsum kettles**	**Gypsum calciners**	**Amulima claciners**	**Mg(OH)2 calciners**	**Nonferrous metals production**	**Sulfuric acid**	**Phosphoric acid**	**Shale oil**	**Phosphorus production**	**Acetylene**	**Waste incinerator**	**Titanium dioxide**	**Coal pyrolysis**	**Pickle liquor**	**Sulfur cleaning**
157 to 346	3.0 to 42.9	7.6 to 82	25 to 135	27 to 115	3.6 to 600	15 to 102	13 to 43	12 to 15	10 to 37	42 to 46	32 to 270	37 to 45	83 to 90	32 to 36	15 to 20
352 to 630	200 to 300	250 to 375	250 to 720	400 to 750	120 to 800	95 to 170	68 to 170	140	520 to 750	100	500 to 700	650	220	750	1292
−4 to +4	−6 to −2	−6 to −2	−10 to +10	Negative	−10 to +10	−36 to 0	−5 to +5	10	−0.3 to +0.3	50	−10 to +10	Negative	Negative	2 psig	–
25 to 35	20 to 40	20 to 40	40 to 50	25 to 50	0.5 to 15	5 to 25	3 to 25	–	None	6	Various	27	–	26	21
9.9 to 53.0	5.0 to 48.0	5.0 to 48.0	100 to 150	2 to 17	0.6 to 30	20 to 200	40 to 100	40	1.5 to 15.0	1.0	1.0 to 4.0	0.9 to 1.2	2.5	13	–
0.006 to 0.08	0.011 to 0.92	0.011 to 0.92	0.04 to 0.10	0.02 to 0.3	0.015 to 0.8	0.5 to 1.0	0.8 to 5.0	0.2	0.02 to 0.04	0.001	0.02 to 0.08	0.009 to 0.012	0.12	0.08	–
20 to 94	5 to 10	5 to 10	14 to 36	13 to 30	7 to 30	8 to 40	8 to 16	14	15	14	14 to 50	14	20	12	14
98.8 to 99.93	95 to 99.94	95 to 99.94	99 to 99.96	98 to 99.5	90 to 99.6	95 to 99.6	90 to 98	99.5	99	99.9	90 to 99	99	95	99	99

Source: By permission from Sickels [71].

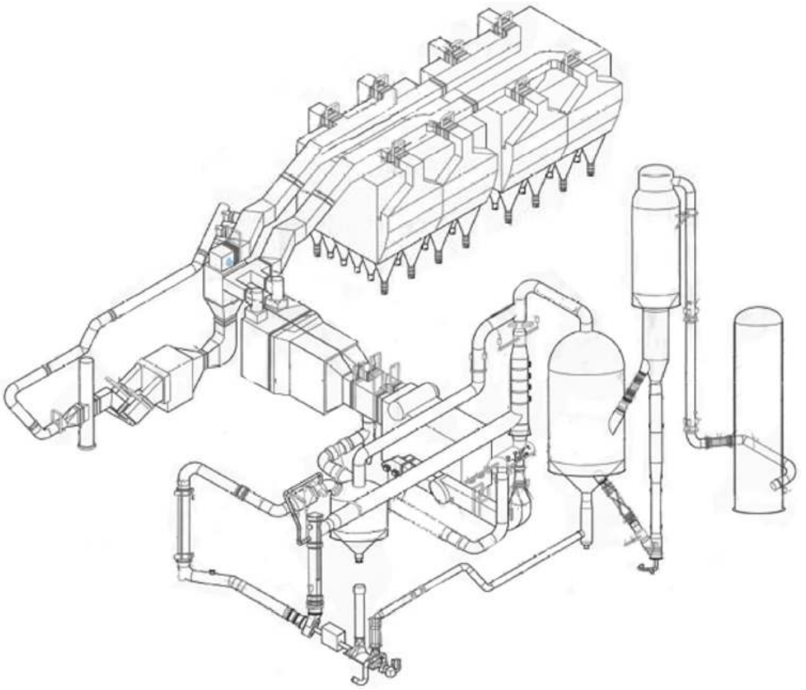


Figure 18.134 Isometric piping connection of the FCC unit, main distillation column, and ESP (source: www.csb.gov).

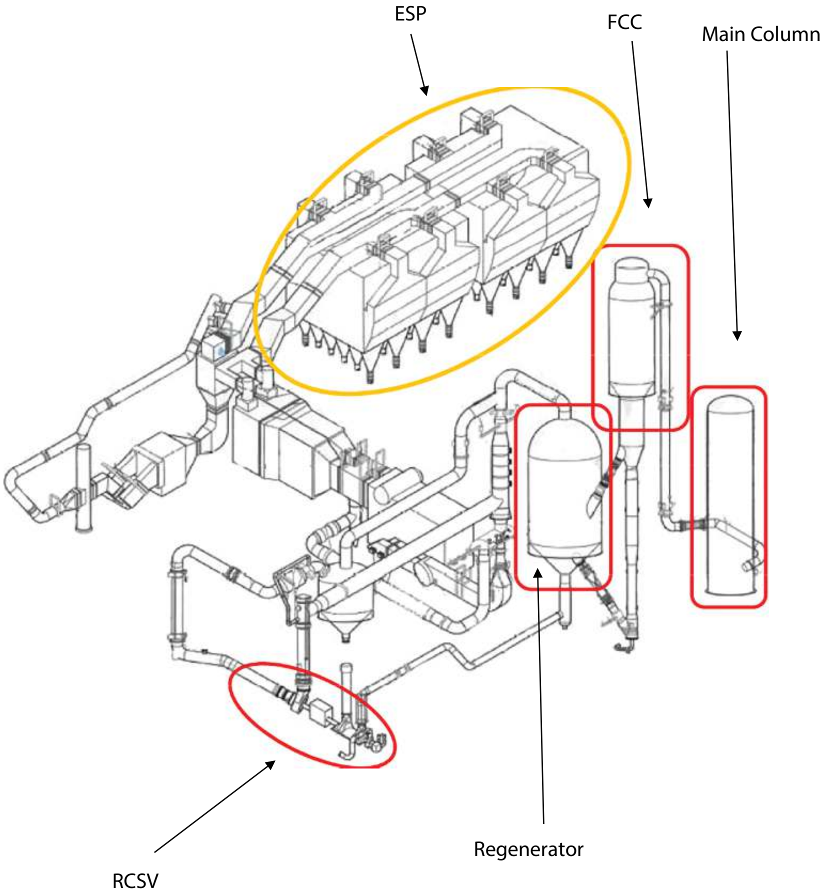


Figure 18.135 Piping isometric circling the FCC, regenerator, main column, electrostatic precipitator, and the regenerated catalyst slide valve (source: www.csb.gov).

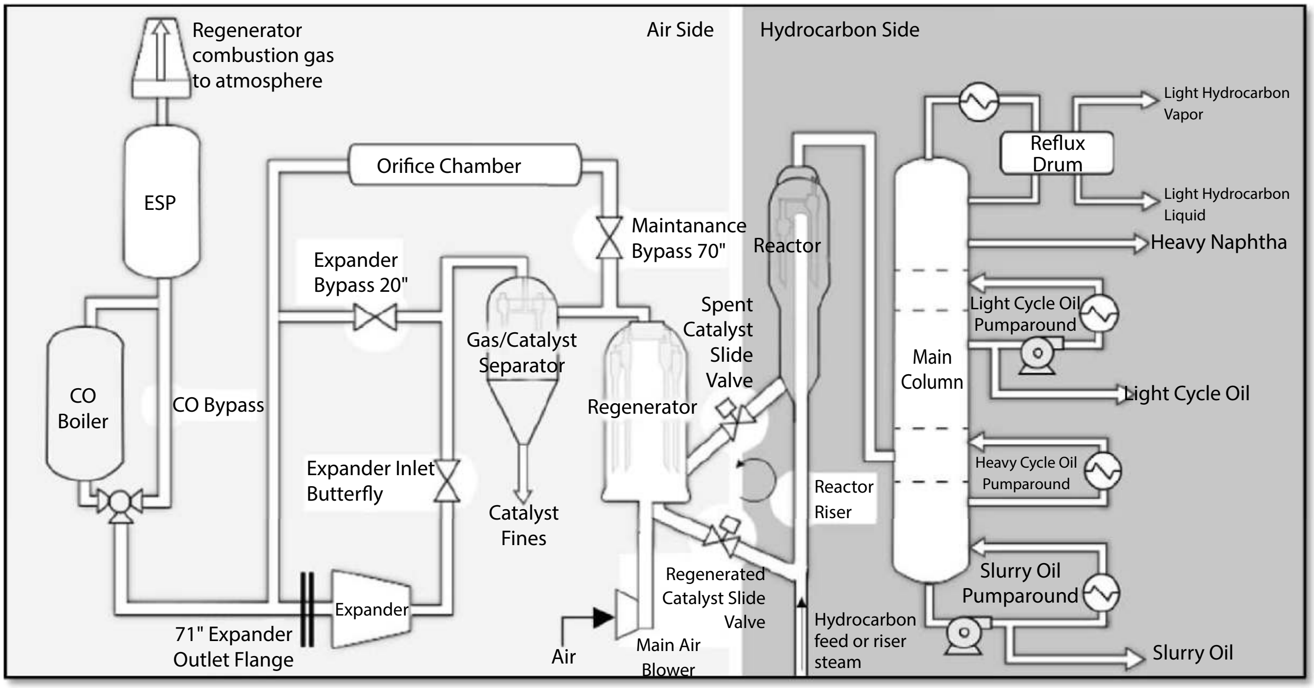


Figure 18.136 Schematic of ExxonMobil Torrance Refinery FCC unit (source: csb.gov).

"spent catalyst". The spent catalyst and cracked hydrocarbon vapor exit the riser and enter the reactor vessel, where most of the catalyst particles are separated from the hydrocarbon vapor. The cracked hydrocarbon vapor then flows to the main column for separation.

The spent catalyst is routed to the reactor standpipe. Within the reactor standpipe is the spent catalyst slide valve (SCSV), which controls the spent catalyst flow into the regenerator. Inside the regenerator, the hot spent catalyst contacts air supplied by the main air blower. The coke deposits that are on the surface of the hot catalyst particles are burned off when in contact with the air in a combustion reaction. The heat of combustion further heats the catalyst particles and this regenerated catalyst enters the regenerator standpipe. The regenerated catalyst slide valve (RCSV) controls the flow of the hot catalyst to the reactor riser where it contacts, heats and vaporizes fresh heavy hydrocarbon feed to begin the cracking reaction.

The cracked hydrocarbon vapors leave the top of the reactor and enter a distillation column referred to as the main column (Figure 18.140). The main column is fed superheated hydrocarbon vapor with no additional heat added to the column during normal operation. Heat is removed from the column to cool and condense the gas feed for separation by removing heat in several loops called pumparounds. In these pumparounds, heat exchangers transfer heat to other process streams in the refinery thereby reducing the temperature of the streams returning to the main column. The main column separates the product from the reactor into light hydrocarbons and heavy naphtha that are further processed to produce gasoline, light cycle oil, and slurry oil.

The gas leaving the top of the regenerator is composed of combustion product gases entrained with catalyst particles. The gas is routed to the gas/catalyst separator as shown in Figure 18.141, where most of the catalyst dust particles are removed from the combustion product gases. The gas, still containing some catalyst dust, flows through the expander, where the expansion of gas is used to partially power the main air blower. Heat is removed from the gas in the carbon monoxide (CO) boiler, and then the gas is routed to the ESP. The ESP collects most of the remaining small catalyst particles from the gas to meet California emissions regulations by using charged plates to attract the fine catalyst particles (Figure 18.142). This operation generates sparks, a potential ignition source inside the ESP.

The FCC unit has two sides: the hydrocarbon and the air sides. The hydrocarbon side includes the reactor and the main column. The air side includes the regenerator and the piping and equipment downstream of the regenerator leading to the ESP (Figure 18.136). The SCSV and the RCSV are used to prevent undesirable mixing of air and

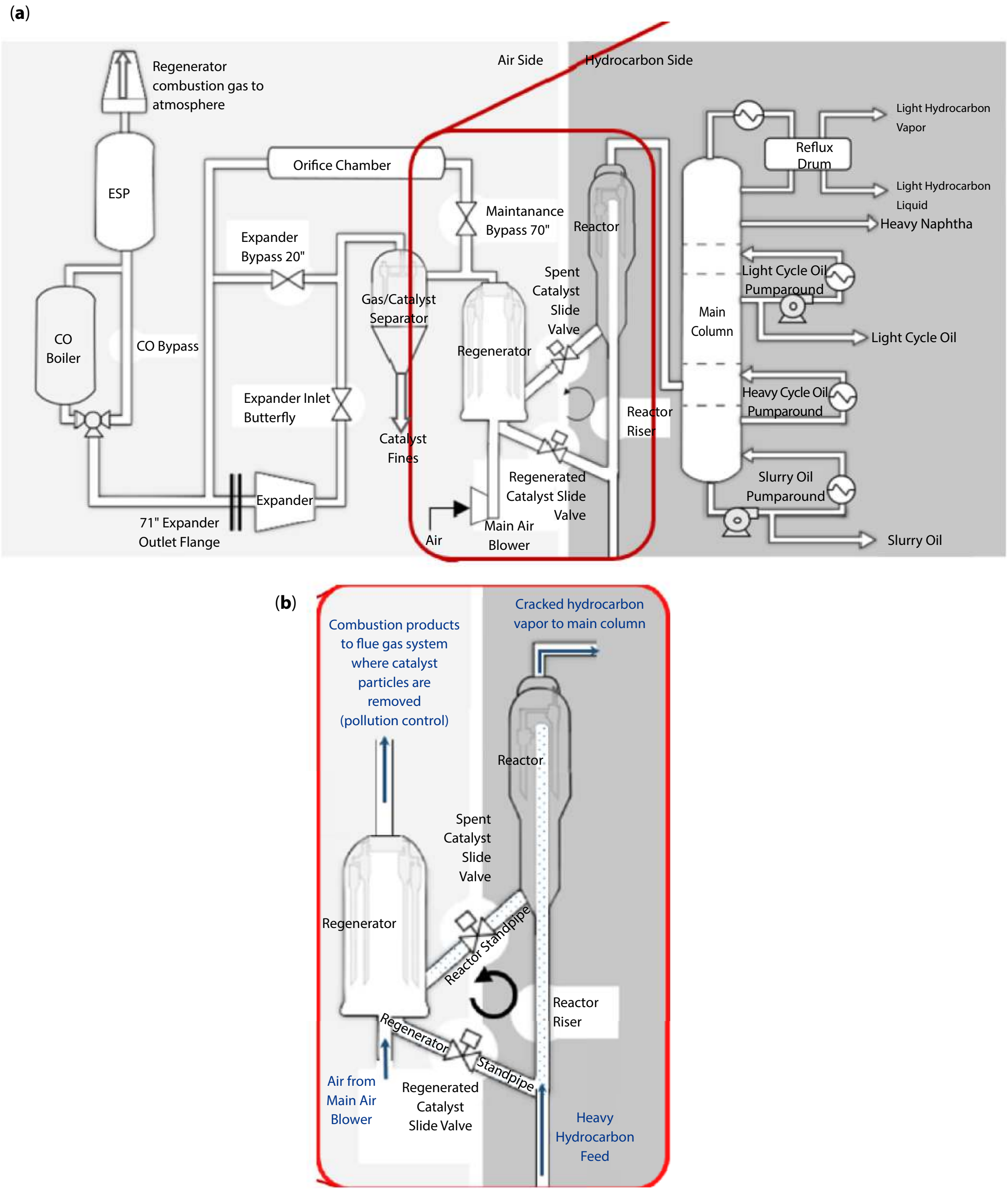


Figure 18.137a Catalyst loop in the FCC unit (source: csb.gov).

hydrocarbons, which is an explosion hazard. During the Safe Park mode of operation (a standby mode of operation that the FCC unit was in on the day of the incident), the two valves isolate the air side and the hydrocarbon side from each other by maintaining a level of catalyst on top of the valves, forming a “plug” that prevents reactor process vapors from entering the regenerator and vice versa (Figure 18.143).

On the day of the incident, a higher than normal pressure in the main column allowed hydrocarbons to enter the reactor with the riser steam flow rate reduced to 7500 lb/h. Leading to the incident, the main column operated at a pressure of about 8.5 psig, roughly double the main column pressure during the 2012 Safe Park. In 2015, the column pressure had increased because a heat exchanger that had an extended operation on the slurry oil pumparound loop was leaking naphtha into the slurry oil (Figures 18.144 and 18.145). The high temperature of the slurry oil vaporized the more volatile naphtha, which increased the pressure of the main column.

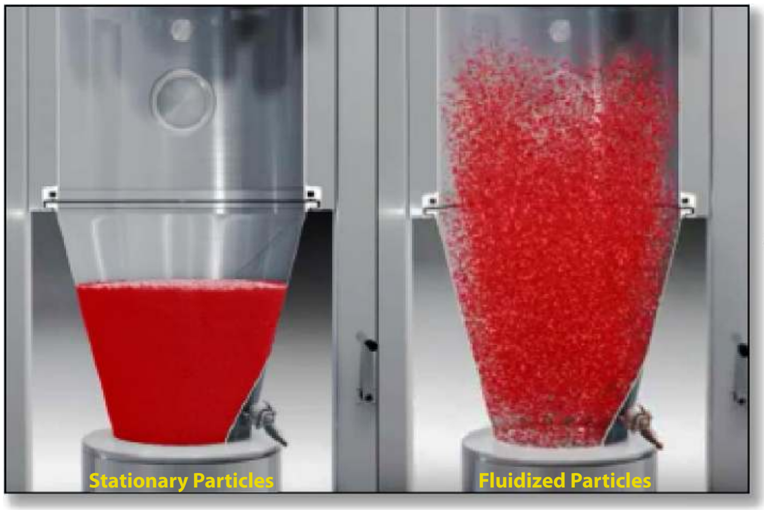


Figure 18.138 Depiction of stationary particles (left) and fluidized particles (right) (source: www.csb.gov).

Figure 18.139 A photograph of the catalyst (source: www.csb.gov).

The heat exchanger that leaked naphtha into the slurry oil pumparound was one of the two heat exchangers that provided heat to a distillation column in another unit. The heat exchangers were designed so that one heat exchanger could be operated while the second heat exchanger remained on standby. The standby exchanger was clean and ready for use when the operating heat exchanger had to be taken offline for cleaning, inspection, and maintenance.

The tubes of the operating heat exchanger were scheduled to be cleaned to remove process buildup and then be visually inspected in September 2013. A block valve that could isolate the heat exchanger from the process, however would not sufficiently close (Figure 18.146), and so the heat exchanger cleaning and inspection could not be completed while the unit was operating. ExxonMobil chose to continue operating the fouled heat exchanger until the next scheduled turnaround in June 2015.

In its Equipment Strategy document, ExxonMobil personnel previously identified that the heat exchanger tubes could corrode and/or erode and leak naphtha into the slurry oil, but they did not identify that such a leak could have negative safety consequences during Safe Park and identified only a minor economic consequence. Furthermore, this specific main column pressurization scenario could be difficult to identify and was not identified by ExxonMobil during hazard analyses.

The SCSV did not maintain the catalyst plug (Figure 18.143). The reduced steam flow no longer prevented hydrocarbons from entering the air side of FCC unit. Hydrocarbons from the main column leaked past the closed eroded spent catalyst sliding valve (SCSV). Hydrocarbons from the reactor flowed into the regenerator in the air side of

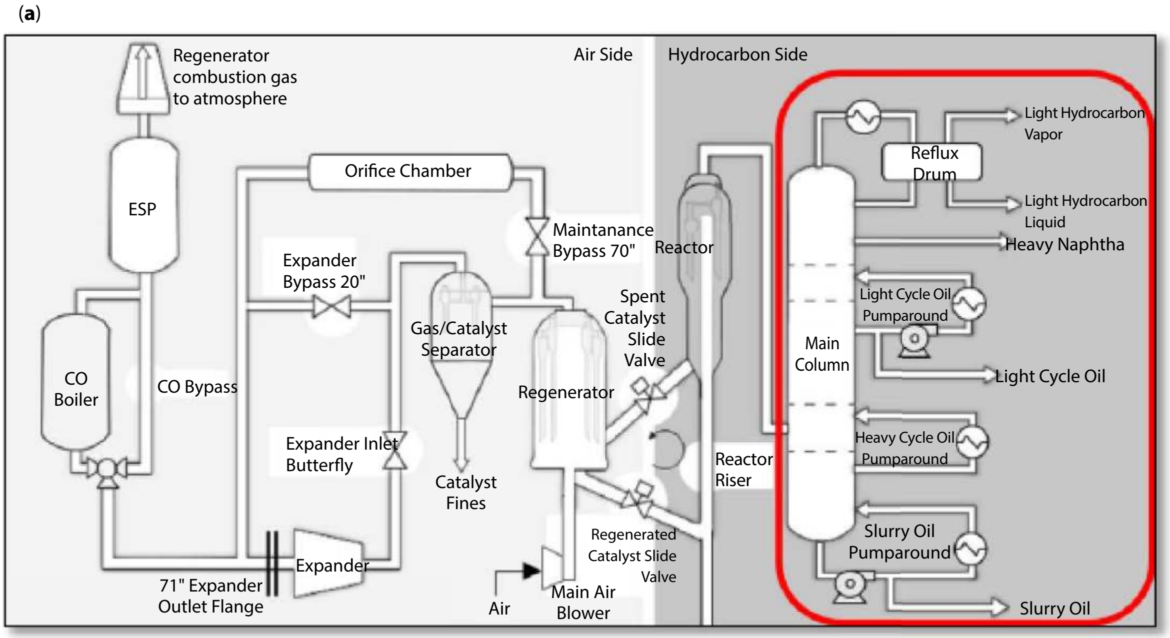


Figure 18.140a FCC unit main column (source: www.csb.gov).

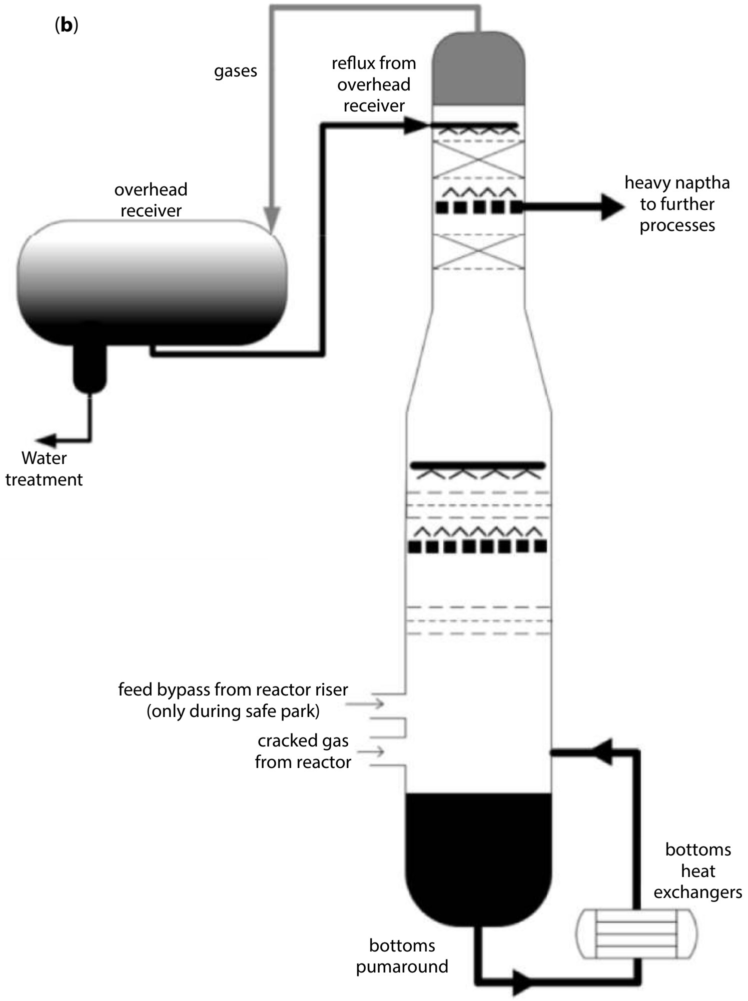


Figure 18.140b Main column operation (www.csb.gov).

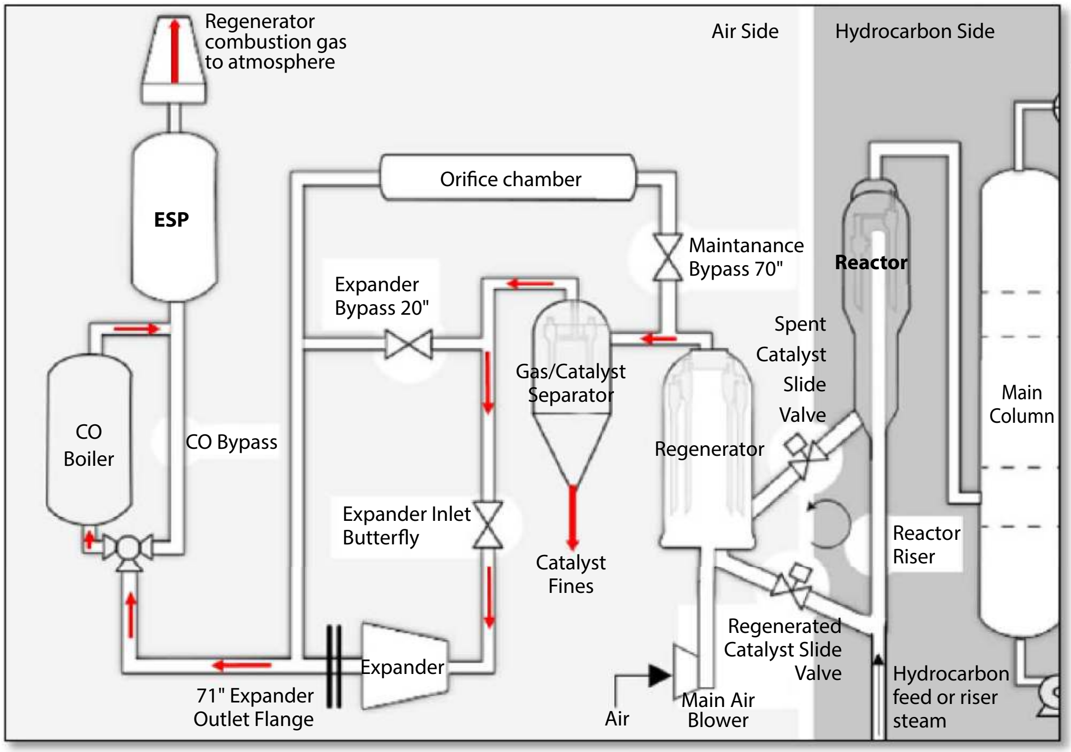


Figure 18.141 Regeneration combustion gas flow to ESP as indicated by the arrow (source: www.csb.gov).

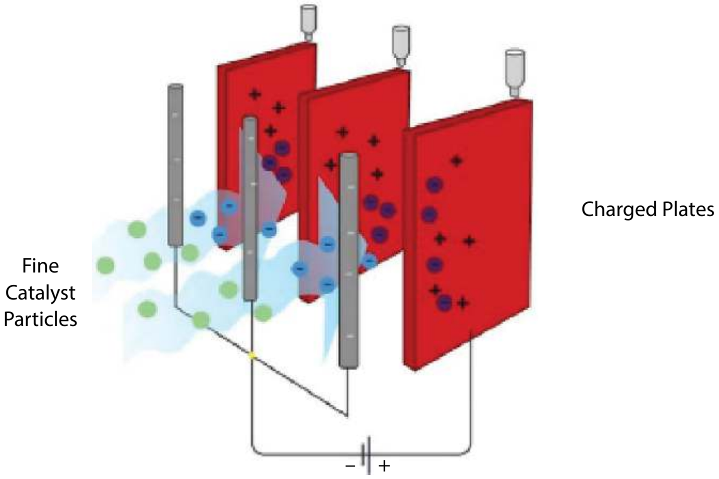


Figure 18.142 The ESP removes catalyst particles using charged plates that produce sparks (as designed) (source: www.csb.gov).

the FCC unit, which in the Safe Park mode of operation was not sufficiently hot to burn (i.e., combust) the hydrocarbons. As a result, flammable hydrocarbons flowed to the ESP, where they mixed with air feed to the ESP from the carbon monoxide (CO) boiler fans, forming a flammable mixture. Sparks within the ESP ignited the flammable mixture causing an explosion (Figure 18.147). Figure 18.148 shows the damage caused to the ESP. A large piece of debris from the explosion fell on scaffolding around two settler tanks, each containing hydrofluoric acid (HF), water, hydrocarbons, and a chemical additive intended to reduce the amount of HF vaporized during a loss of containment event (Figure 18.149). Pure HF is a highly toxic chemical that can seriously injure or cause death at a concentration of 30 ppm. Figures 18.150 and 18.151 show photographs of the damaged plant, respectively.

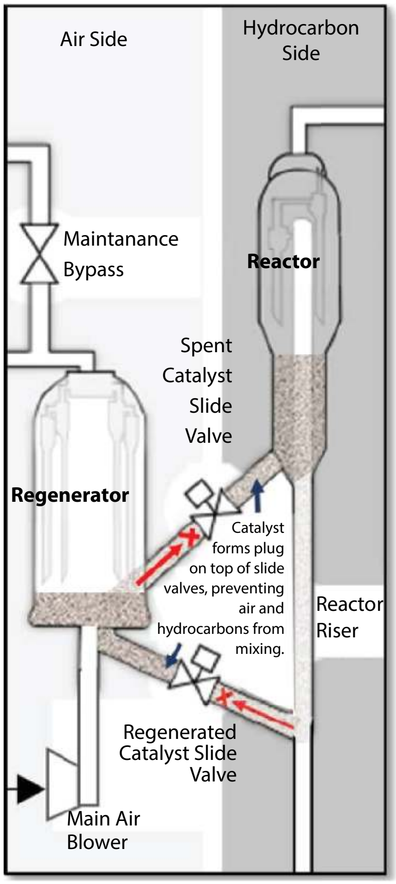


Figure 18.143 During Safe Park, a catalyst level on top of the SCSV and RCSV is intended to prevent hydrocarbons and air from mixing (source: www.csb.gov).

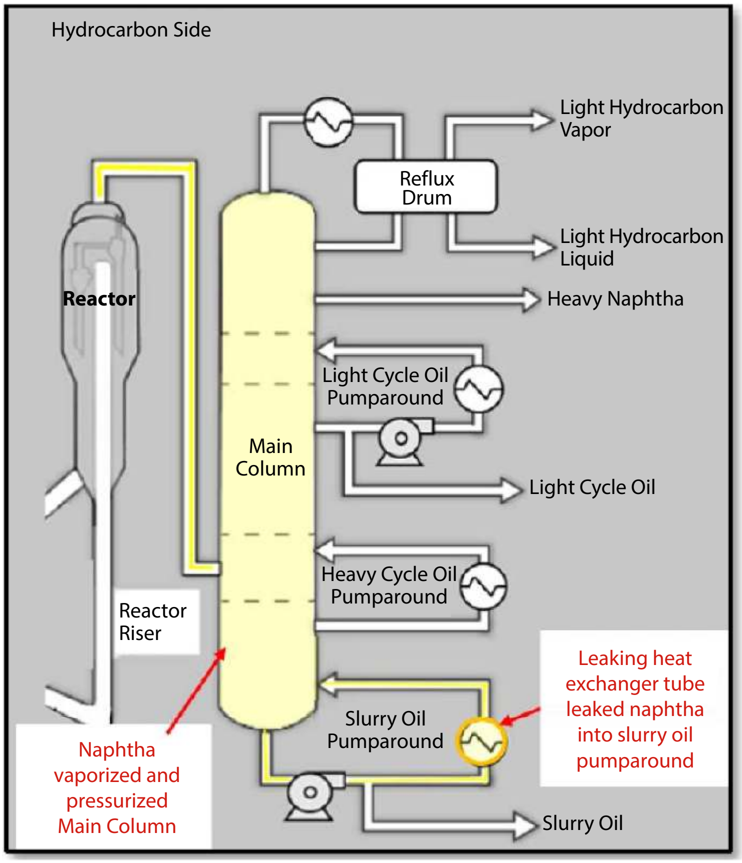


Figure 18.144 Leaking heat exchanger tubes allow light naphtha to enter main column (source: www.csb.gov).

Figure 18.145 Post-incident photos of heat exchanger tube bundle that leaked naphtha into main column slurry oil pumparound (source: www.csb.gov).

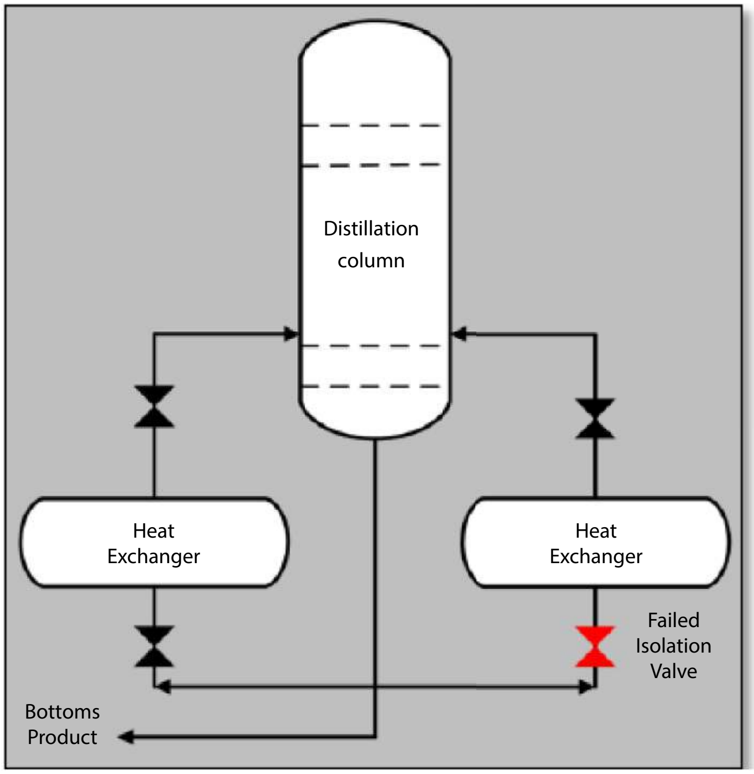


Figure 18.146 Heat exchanger shell-side path showing isolation block valve that would not close (source: www.csb.gov).

Key Factors That Contributed to a Flammable Mixture Accumulating Inside of the Electrostatic Precipitator (ESP)

The U.S. Chemical Safety and Hazard Investigation Board (CSB) identified the casual factors that resulted in the explosion of the electrostatic precipitator (ESP) as the following:

1. ExxonMobil relied on indirect operating parameters to measure critical safeguards for the Safe Park mode of operation. They relied upon operating parameters that did not signify that the FCC unit was in a dangerous condition leading to the incident. Additionally, ExxonMobil had not developed a Safe Park procedure for how to safely operate within specified safe operating limits.
2. In 2015, ExxonMobil relied on a Variance that had been developed in 2012 without verifying that the safeguards specified in the Variance were sufficient.

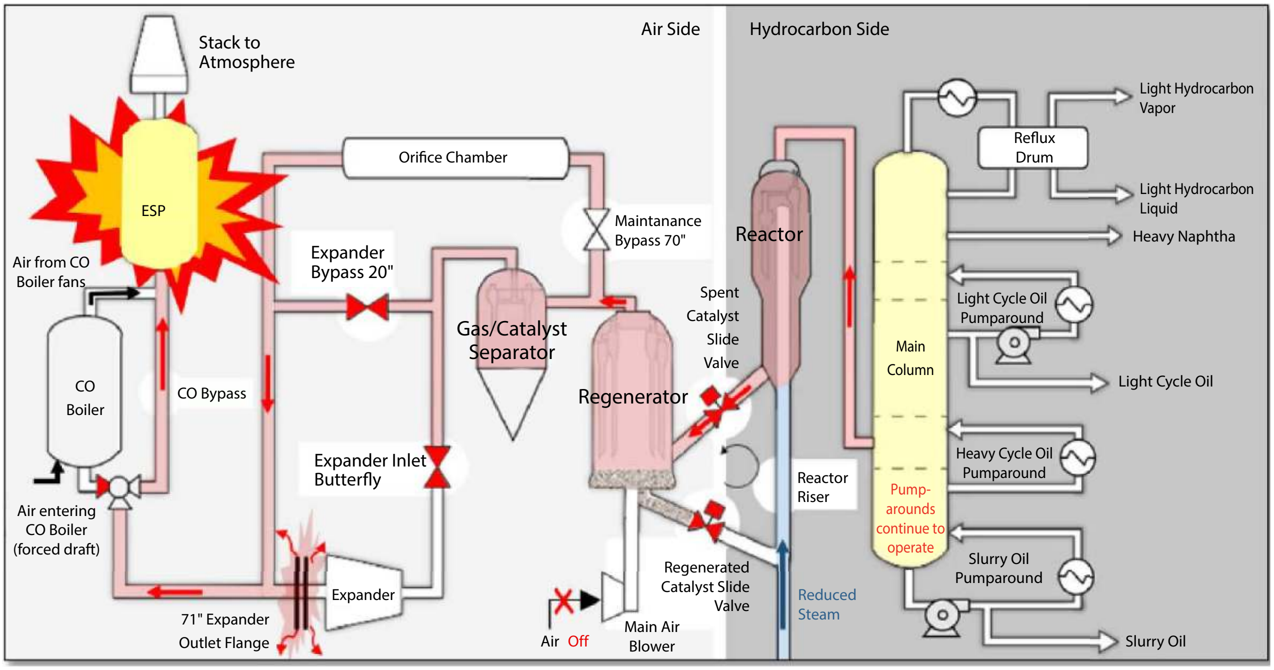


Figure 18.147 When steam flow rate was reduced through the reactor riser, reactor pressure could no longer prevent hydrocarbon backflow from the main column. Hydrocarbons leaked through closed SCSV, through FCC unit air side piping and equipment, and to the energized ESP. A hydrocarbon/air mixture ignited inside the ESP (source: www.csb.gov).

Figure 18.148 Damage to the ESP from the explosion. White catalyst dust from the ESP is visible on the ground in left photo (source: www. csb.gov).

3. Erosion damage that had developed over 6 years of operation likely compromised the spent catalyst slide valve (SCSV), and it could not maintain a catalyst barrier while the FCC unit was in Safe Park.
4. Steam flow to the reactor had been reduced, likely causing a pressure deviation that allowed hydrocarbons to enter the flue gas system.
5. A leaking heat exchanger in the slurry oil pumparound allowed light hydrocarbons to enter and pressurize the main column to a higher than typical pressure.
6. ExxonMobil did not shut down the FCC unit when it was identified that the SCSV leaked and had not established a catalyst barrier.
7. The expander could not be effectively isolated while the unit was in Safe Park. ExxonMobil opened process equipment without conforming to refinery standards.
8. The ESP remained energized when hydrocarbons entered the flue gas system, providing an ignition source to trigger the explosion.

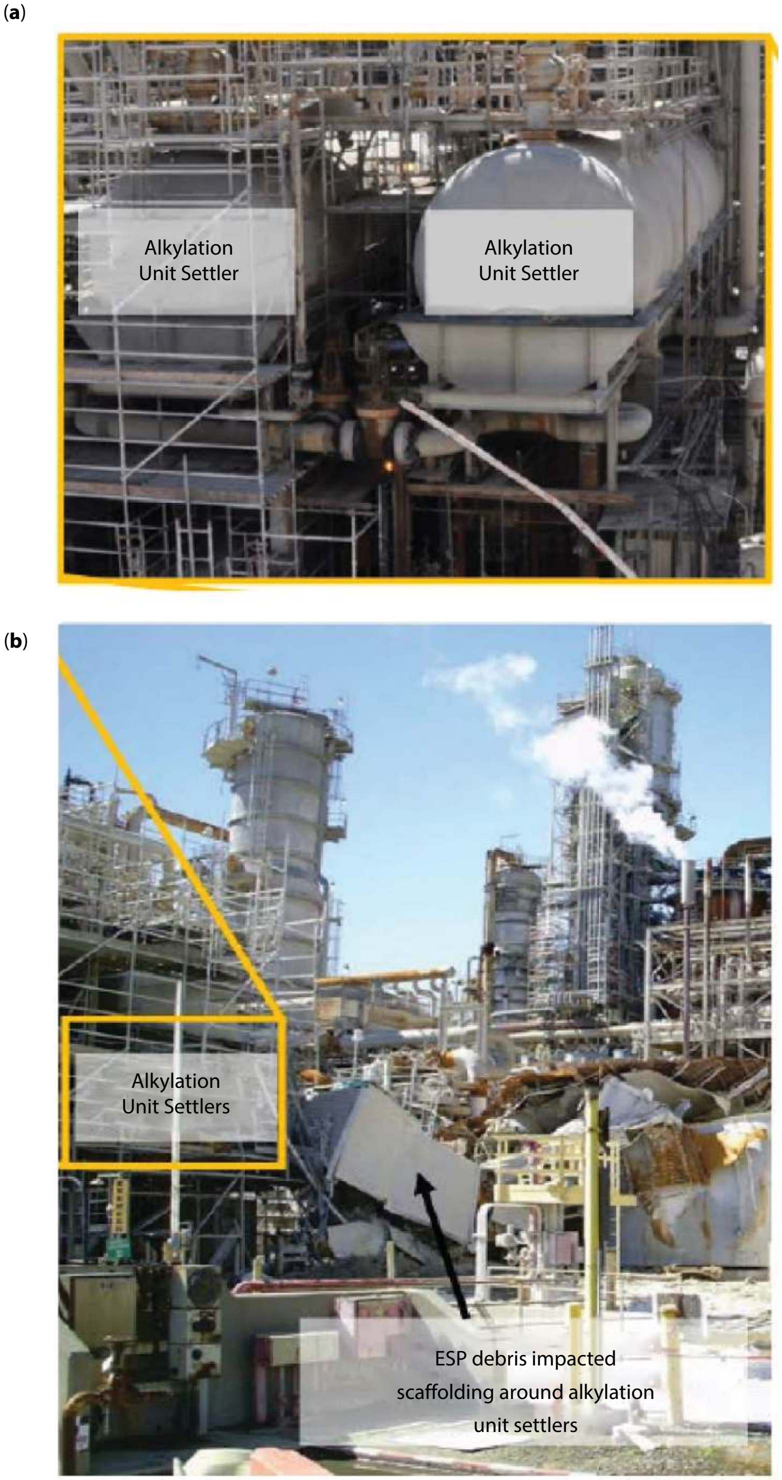


Figure 18.149a Explosion debris landed in close proximity to alkylation unit settler tanks containing hydrofluoric acid (HF), water, hydrocarbons, and a chemical additive intended to reduce the amount of HF vaporized during a loss of containment event (source: www.csb.gov).

ExxonMobil performed a damage mechanism hazard review of the SCSV. The hazard review correctly identified that erosion was a damage mechanism that affected the SCSV. However, the company identified only a financial consequence of the SCSV failure by erosion. Despite the SCSV classification as a safety critical device, the company did not identify a safety and health consequence. The CSB found that ExxonMobil considered only normal operating

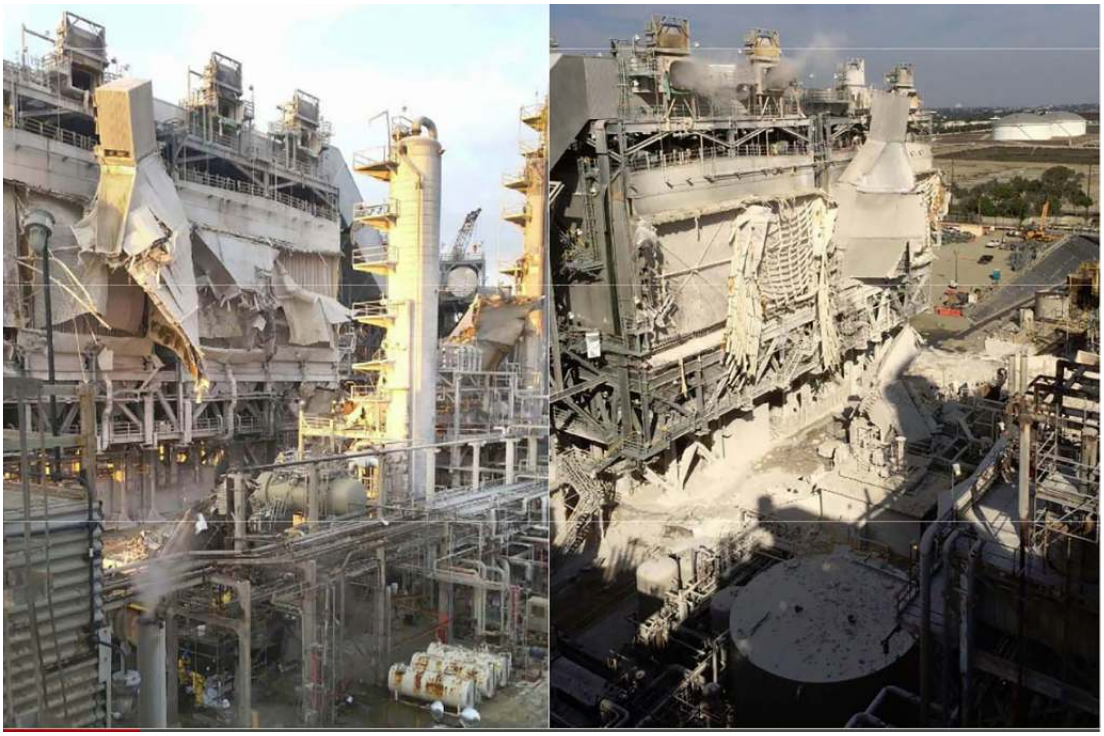


Figure 18.150 A photograph showing the ESP, distillation column, and ancillaries (source: www.csb.gov).

Figure 18.151 A photograph showing the ESP and damaged piping (source: www.csb.gov).

conditions when identifying consequences of failure as the company did not consider that an eroded SCSV may not maintain a catalyst plug while the FCC unit was in Safe Park.

Key Findings Identified in the CSB Investigation

The principal findings from the CSB investigation are:

- Implementation of ExxonMobil Operations Integrity Management System (OIMS)

- Process Hazard Analysis (PHA)
- Mechanical Integrity
- Worker Participation
- Process Safety Regulatory Gaps
- Community Impacts

Process Hazard Analysis: The CSB found that ExxonMobil failed to adequately consider hydrocarbons reaching the ESP.

- Process hazard analysis only considered carbon monoxide (CO) entering the ESP. As a result, detectors used to shut down ESP were only calibrated to read CO.
- Investigating team could not determine the rationale behind the ExxonMobil PHA process.
- Spent catalyst slide valve is not maintaining catalyst level.
- Hydrocarbons in the main column.
- No analysis of steam flow rate was needed in the column.

Mechanical Integrity

- Pressure transmitter in the main column.
- Expander
- CO gas detectors
- Heat exchanger leaking naphtha
- Valves attached to these heat exchangers.

Potential Health Effects of Hydrofluoric Acid and Modified Hydrofluoric Acid

Hydrofluoric acid (HF) is a toxic chemical and poses severe hazard to the population when a release occurs.

- Causes severe damaged to skin, respiratory system and bones after exposure can lead to death.
- Large release could impact hundreds of thousands of residents.

Community

- A large quantity of catalyst dust was released due to the ESP explosion.
- Catalyst dusts fell in the nearby communities as far as a mile away.

Lack of Hierarchy of Controls (HOC) Analysis

- Could have prevented incident through utilization of passive barriers.
  - Shutting all valves leading to the ESP or
  - A blind at the top of the main column.
- HOC principles could have been applied.
  - Design spent catalyst slide valve (SCSV)
  - Removing hydrocarbons (HC) from main column (MC)
  - Evaluate the use of modified hydrofluoric acid (HF).

Process Safety Management (PSM)

A process safety management system (PSM) approach focused on the prevention and mitigation of catastrophic releases of chemicals or energy from a process associated with the facility.

Elements of PSM include:

- Process Hazard Analysis (PHA)
- Operating Procedures

- Management of Change (MOC)
- Employee participation.

However, some PSM failures would not have been identified under the current California PSM regulation. This is because:

- Certain PSM elements fail to require an assessment of their adequacy of completion
- Current PSM regulation also lacks key process safety requirements.

Process safety management systems are reviewed in Chapter 23, Volume 4 of these volume series.

Additional Findings by the CSB are:

- Organizational failures
- ESP Siting
- Non routine operating conditions
- Safety critical equipment management

The US Chemical Safety and Hazard Investigation Board (CSB) Board Key Lessons

The CSB investigated this incident and pertinent lessons from their investigation are as follows:

1. It is essential to identify and define safe operating limits for all modes of operation measure process conditions and parameters that can verify the operation of the process relative to those safe operating limits. When a facility relies on operating parameters that only indirectly provide information on critical process parameters, it can lead to the inability to identify when a process is in an unsafe condition.
2. When implementing a deviation from an existing procedure, it is critical that the company conduct a management of change to (MOC) among other requirements, verify and authorize the technical basis, the implementation time period, and identify any new or affected hazards and associated mitigation strategies. If the procedure deviation is saved for future use, before implementing the procedure ExxonMobil should verify that the underlying conditions, activities, and technical assumptions that were the basis for the initial authorization are in place and are still valid.
3. It is essential to schedule and perform maintenance of safety critical equipment so that the equipment is available to perform its safety critical function.
4. It is important to consider all modes of operation, including non routine operations such as unit standby when performing process hazard analyses. Incident scenarios could be possible during non routine modes of operation that may not have been considered when analyzing process hazards for normal, continuous operation.
5. Companies should develop operating procedures for all modes operation including unit standby that detail safe operating limits, consequences of deviating from those limits, and specified actions to implement in the event the process deviates outside of its safe operating limits.
6. Robust management of change practices are needed when making changes to procedures. Similar to process hazards analysis (PHAs), conducting MOC as a multidisciplinary group composed of individuals with different areas of expertise can assist in identifying hazards introduced by the procedure change.
7. Control valves typically should not be used as block valves because fluid flow through a partially open control valve can cause damage to the valve that can limit its ability to fully seal.
8. Uncombusted hydrocarbons that are not accompanied by carbon monoxide have the potential to reach FCC unit electrostatic precipitators (ESPs). Refining companies should evaluate their FCC units to determine whether there are sufficient safeguards to prevent an ESP hydrocarbon explosion.

Conclusions

The CSB concluded that the incident of explosion at Torrance refinery was preventable and the weakness in the ExxonMobil process safety management program led to a hydrocarbon backflow in the FCC unit and ignition in the

ESP. The company did not develop a procedure specifically for operating in Safe Park that established safe operating limits and the process conditions that required unit shutdown. Furthermore, the ExxonMobil did not adequately define the function of its safety critical equipment while in Safe Park, and did not ensure the safety critical equipment could perform its safety critical function. ExxonMobil did not sufficiently perform risk analyses to identify the adequacy of its Safe Park safeguards and effective safeguards were not established to prevent the incident.

The CSB identified several process safety design weaknesses such as the piping and equipment between the regenerator and ESP and these were not configured with instrumentation to detect hydrocarbons flowing toward the ESP. The inability to detect hydrocarbons in piping and equipment leading to a potential hazard may be a process safety deficiency. Furthermore, the spent catalyst slide valve could not reliably isolate the hydrocarbon side and air side of the FCC unit from one another. ExxonMobil relied on the SCSV as a safety critical block valve while in Safe Park, but the SCSV was designed to be a control valve and could not adequately seal.

The CSB Recommendations

1.	A Variance to a safety policy or procedure requires robust analysis of the proposed safeguards prior to its approval and implementation. To ensure the proposed methodology described in the Variance is safe and the proposed safeguards are sufficiently robust, revise corporate and U.S. refinery standard(s) to require that a multidisciplinary team reviews the Variance before it is sent to management for approval. The Variance multidisciplinary team should be: 1. The developer of the Variance 2. A technical process representative (e.g., process engineer for the applicable unit). 3. An hourly operations representative (e.g., experienced operator in the applicable unit) 4. A health and safety representative The role of the team is to formally meet to review, discuss, and analyze the proposed Variance, and adjust the safety measures as needed to ensure a safe operation. Where the expert team members do not come to a consensus that the Variance measures can result in a safe operation, then the proposed work be escalated to a higher management lever for final approval.
2.	ExxonMobil did not have an operating procedure for operating the FCC unit in its Safe Park mode of operation. At all ExxonMobil U.S. refineries, a program should be developed to ensure operating procedures are written and available for each mode of operation such as unit standby for all ExxonMobil U.S. refinery FCC units. The program must develop and train operators or any new procedure.
3.	The spent catalyst slide valve, specified as a safety critical device for normal operation, could not perform its safety critical function of preventing air and hydrocarbons from mixing while the FCC unit was in its "Safe Park" mode of operation. Furthermore, ExxonMobil Torrance did not operate the FCC unit as if the reactor steam was a safety critical safeguard. Identification is required of all safety critical equipment and consequence of failure for each mode of operation and to ensure that safety critical devices can successfully function when needed. Develop and implement a policy that requires all U.S. ExxonMobil refineries to: 1. Specify each safety critical device's safety function. 2. Identify the consequences of failure of each safety critical device. 3. Specify the testing strategy used to verify whether the safety critical device can function as intended to perform its required safety function. 4. Maintain target availability (e.g. safe operating life) for each safety critical device through inspection and maintenance. Require that items 1–4 consider each mode of operation, including but not limited to normal operation, start up, shut down and "Safe Park" modes of operation.

4.	ExxonMobil extended the maintenance interval of the spent catalyst slide valve and the inspection interval of the pumparound heat exchanger without analyzing whether the extended operation lowered their availability (by operating them beyond their predicted safe operating life) and could result in negative safety consequences. In the event safety-critical equipment is operated beyond its inspection and/or maintenance interval (e.g., extended turnaround interval), require all ExxonMobil U.S. refineries to perform a risk evaluation (e.g., MOC or risk assessment) to identify the safety consequences of the extended operation. Require each mode of operation, including but not limited to normal operation, start up, shut down and "Safe Park" modes of operation is evaluated during the risk evaluation.
5.	Electrostatic precipitators create potential ignition sources during normal operation and have historically caused explosions within the refining industry. At all U.S. ExxonMobil refineries, require a siting risk analysis be performed of all electrostatic precipitators and implement appropriate safeguards to minimize the consequences of an electrostatic precipitator explosion.

Readers can visit the US Chemical Safety Board website to view the animated video of ExxonMobil refinery in Torrance California (www.csb.gov/exxonmobil-refinery-explosion).

Nomenclature

a = Specific surface area, sq ft/cu ft
a_e = Acceleration due to gravity, 32.2 ft/s^2 or 9.8 m/s^2
A = Area of segment of a circle, sq ft
or, A = Cross-sectional flow area, sq ft
A_b = Cross-sectional area at bottom of vessel occupied by continuous aqueous phase, sq ft
A_c = Cyclone inlet area = $B_e H_c$ for cyclone with rectangular inlet, sq ft
A_I = Area of interface, assumes flat horizontal, sq ft
A_H = Cross-sectional area allocated to heavy phase, sq ft
A_L = Cross-sectional area allocated to light phase, sq ft
A_p = Area of particle projected on plane normal to direction of flow or motion, sq ft
A_t = Cross-sectional area at top of vessel occupied by continuous hydrocarbon phase, sq ft ACFS
= Actual flow at conditions, cu ft/s
b_1 = Constant given in table
c = Volume fraction solids
C = Overall drag coefficient, dimensionless
D = Diameter of vessel, ft
D_b = See D_p, min
D_c = Cyclone diameter, ft
D_e = Cyclone gas exit duct diameter, ft
D_H = Hydraulic diameter, ft = 4 (flow area for phase in question/wetted perimeter); also, D_H in decanter design represents diameter for heavy phase, ft
D_L = Diameter for light phase, ft
Dp = Diameter of particle, ft or equivalent diameter of spherical particle, ft
D_{p-min} = Minimum diameter of particle that is completely collected, ft
D'_p = Diameter of particle, in. or mm
d = Droplet diameter, ft
f = Factor relating average velocity to maximum velocity
f_c = Friction factor, dimensionless
F = Flow rate of one phase, GPM
F_{aq} = Aqueous phase flow rate, GPM
F_{ev} = Cyclone friction loss, expressed as number of Cyclone inlet velocity heads, based on A_c
F_d = Drag or resistance to motion of body in fluid, poundals

F_{hc} = Hydrocarbon phase flow rate, GPM
F_t = Total flow rate of both phases, GPM
$g = g_c = g_L$ = Acceleration due to gravity, 32.2 ft/(s) (s)
h = Distance from center to given chord of a vessel, ft
h_b = Height of continuous aqueous phase in the bottom of the vessel, in.
h_c = Height of a segment of a circle, in.
h_t = Height of continuous hydrocarbon phase in the top of the vessel, in.
h_{vi} = Cyclone inlet velocity head, in. water
H = Height of a segment of a circle, ft
H_c = Height of rectangular cyclone inlet duct, ft
H_D = Height of dispersion band, ft
I = Width of interface, ft
$k = K$ = Empirical proportionally constant for cyclone pressure drop or friction loss, dimensionless
K' = Constant for stationary vane separators, based on design
K_m = Stokes-Cunningham correction factor, dimensionless
K_{me} = Proportionality factor in Stokes-Cunningham correction factor, dimensionless
k = Constant for wire mesh separators
1 = Wire mesh thickness, ft
L = Length of vessel from hydrocarbon inlet to hydrocarbon outlet, or length of decanter, ft
L_I = Liquid entering Webre separator, lbs per minute per square foot of inlet pipe cross-section
L_v = Entrainment from Webre unit, lb liquid per minute per square foot of inlet pipe cross-section
m = Exponent given by equations
m_p = Mass of particle, lb mass
n = Constant given in table
N_{Re} = Reynolds number, dimensionless (use (Re) consistent units)
N_t = Number of turns made by gas stream in a cyclone separator
ΔP = Pressure drop, lbs/sq in.
Δp = Pressure drop, in. water
ΔP_D = Pressure drop, no entrainment, in. water
ΔP_L = Pressure drop due to liquid load, in. water
ΔP_T = Pressure drop, total across wet pad, in. water
Q_D = Dispersed phase volumetric flow rate, cu ft/s
Q_H = Volumetric flow rate, heavy phase, cu ft/s
Q_L = Volumetric flow rate, light phase, cu ft/s
r = Vessel radius, ft
SpGr = Specific gravity of continuous phase at flow conditions
$SpGr_p$ = Specific gravity of settling particle at flow conditions
$\Delta SpGr$ = Difference in specific gravity of the particle and the surrounding fluid
t_{ave} = Average residence time based on liquid flow rate and vessel volume, min
t_{min} = Minimum residence time to allow particles to settle based on Stokes Law, min
u = Relative velocity between particle and main body of fluid, ft/s
u_t = Terminal settling velocity determined by Stokes Law, of particle under action of gravity, ft/s
u_{ts} = Terminal settling velocity as calculated from Stokes Law, ft/s
$v = V_t$ = Terminal settling velocity, in./min
v_a = Average velocity of gas, ft/s
v_{ag} = Terminal settling velocity of hydrocarbon droplets in aqueous phase in bottom of vessel, in./min
v_c = Velocity down flow channel for continuous phase, ft/s
v_d = Terminal settling velocity of a droplet, ft/s
v_{hc} = Terminal settling velocity of aqueous droplets in hydrocarbon phase in top of vessel, in./min
v_t = Terminal settling velocity of particle under action of gravity, ft/s
v_{ts} = Terminal settling velocity of particle as calculated from Stokes Law, ft/s

V = Velocity of gas or vapor entering, ft/min
$V_{(separator)}$ = Separator vapor velocity evaluated for the gas or vapor at flowing conditions, ft/s
V' = Vapor velocity entering unit, lbs, per minute per square foot of inlet pipe cross section
V_a = Maximum allowable vapor velocity across inlet face of mesh calculated by relation, ft/s
V_{act} = Actual operating superficial gas velocity, ft/sec or ft/min, for wire mesh pad
V_D = Design vapor velocity (or selected design value), ft/s
V_c = Cyclone inlet velocity, average, based on area A'_c ft/s
V_{max} = Calculated maximum allowable superficial gas velocity, ft/s, or ft/min wire mesh pad
V_s = Superficial gas velocity, ft/s
V_{sa} = Separator vapor velocity evaluated for air-water system, ft/s
V_{set} = Active volume of settler occupied by one of the phases, cu ft
V_t = Settling velocity for single spherical particle, ft/s or m/s
V_{ts} = Settling velocity for hindered uniform spherical particle, ft/s or m/s
W_i = Width of rectangular cone inlet duct, ft
z_h = Heavy phase outlet dimensions of decanter measured from horizontal bottom, shown in Figure 6.12
z_i = Interface of decanter liquids measured from bottom, Figure 6.12
z_l = Light phase outlet measured from bottom of decanter, Figure 6.12

Subscripts

L or 1 = Light phase
H or h = Heavy phase
C or c = Continuous phase
D or d = Dispersed phase
l = Liquid
v = vapor or gas

Greek Symbols

ε = Void fraction of wire mesh, dimensionless
η = Fraction of dispersoid in swept volume collected on target
θ = Factor for establishing type of flow for decanters, Reference [32]
μ = Viscosity of surrounding fluid, cP, except where it is lb/(ft-s)
μ_c = Viscosity of continuous phase, lb/(ft) (s)
μ_H = Viscosity of heavy phase, lb/(ft) (s)
μ_v = Viscosity of fluid, cP
μ_L = Viscosity of light phase, lb/ft sec
μ = Fluid viscosity, (lb mass)/(ft) (s) = centipoise/1488
μ_m = Milli-micron = 0.001 millimeter
π = 3.1416
$\rho = \rho_d$ = Fluid density, or density of fluid in droplet, lb mass/cu ft
ρ_c = Density of fluid continuous phase, lb/cu ft
ρ_f = Density of fluid, lb/ft^3 or kg/m^3
ρ_L = Liquid density, lb/cuft
ρ_d = Density of fluid continuous phase, lb/cu ft
ρ_L = Density of light phase fluid, lb/cu ft
ρ_p = Density of particle, lb/cu ft
$\rho_s = \rho_s$ = True density of particle, lb mass/cu ft
ρ_v = Vapor density, lb/cu ft

References

1. Fair, J. R., "A Half-Century of Progress in Separations Technology", Chemical Processing, Mid-March, 1988.
2. Friedlander, S. K, L. Silverman, P. Drinker, and M. W. First, *Handbook on Air Cleaning Particulate Removal*, United States Atomic Energy Commission, 1952, Washington, D. C.
3. Svarosky, J. K., "Sedimentation , Centrifugation and Flotation", *Chem. Eng*., V. 86, July 16, 1979, p. 93.
4. Alden, J. L., *Design of Industrial Exhaust Systems*, 2nd Ed. Industrial Press, 1940, New York, N.Y.
5. Perry, John H., Ed. *Chemical Engineer's Handbook*, 3rd Ed., "Dust and Mist Collection", by C. E. Lapple, 1950, McGraw-Hill Book Co., Inc.
6. Public Health Bulletin No. 217, *The Determination and Control of Industrial Dust*, U. S. Public Service, 1935.
7. Bulletin, Sales Book Sheet, DC-271, American Air Filter Co., 1953, Louisville, Ky.
8. Perry, R. H. and D. Green, *Chemical Engineers Handbook*, 6th Ed., McGraw-Hill Book Co., 1984, p. 20–82, 5–67, 21–65, 5–66.
9. Sylvan, S., *Range of Particle Sizes, Concentration and Collector Performance*, The American Air Filter Co. 1952.
10. Kane, John M., *Operation, Application and Effectiveness of Dust Collection Equipment, Heating and Ventilating*, August 1952.
11. Vandegrift, A. E., Shannon, L. D., and P. G. Gorman, "Controlliing Fine Particles", Chem. Eng./Deskbook, Vol. 80, No. 14, p. 107, 1973.
12. Montrose, C. F., "Entrainment Separation", *Chem. Eng*., Oct. 1953.
13. Carpenter, C.R., "Calculate Settling Velocities for Unrestricted Particles or Hindered Settling", *Chem. Eng*., V. 90, No. 23, 1983, p. 227.
14. American Petroleum Institute, Division of Refining, Manual on Disposal of Refinery Wastes, Volume on Liquid Wastes, 1st ed. New York, 1969.
15. Applegate, R., L. Gurnari and W. Willerdorf, "How to Select a Lamella Clarifier", *Pollution Engineering*, Jan. 1989, p. 112.
16. Thomson, S. L, "Data Improves Separator Design", *Hydrocarbon Processing*, V. 52, No. 10, 1973, p. 81.
17. Miranda, J. G., "Designing Parallel-Plates Separators", *Chem. Eng*., V. 84, No. 3, 1977, p. 105.
18. Fitch, B., "Gravity Sedimentation Operations", *Encyclopedia of Chemical Processing and Design*, J.J. McKetta, Exec. Ed., M. Dekker, Inc., Vol. 25, 1987, New York, p. 1.
19. Abernathy, M. W. "Design of Gravity Settlers", *Encyclopedia of Chemical Processing and Design*, J. J. McKetta, Exec. Ed., M. Dekker, Inc., Vol. 25, 1987, New York, p. 77. Also, Hydrocarbon Processing, Sept. 1977, p. 199.
20. Hooper, W. B. and L. J. Jacobs, *Handbook of Separation Techniques for Chemical Engineers*, P. A. Schweitzer, Ed. in Chief, McGraw-Hill, 1979, p. 1–345.
21. Ludwig, E. E., Applied Process Design For Chemical and Petrochemical Plants, Vol. 1, 3rd Gulf Publishing Co., Houston, TX, 1995.
22. Happel, J. and D. Jordan, Chemical Process Economics, 2nd Ed., Marcel Dekker, New York, 1975, p. 483.
23. Happel, J., *Chemical Process Economics*, Wiley, 1958, p. 251.
24. Sigales, B., "How to Design Settling Drums", *Chem. Eng*., June 23, 1975, p. 141–144.
25. Schweitzer, P. A., *Handbook of Separation Techniques for Chemical Engineers*, McGraw-Hill Book Co., 1979, p. 1–346.
26. Carpenter, C. L., D. Sc. Ch.E. Dissertation, Polytechnic Institute of Brooklyn, 1951.
27. Bulletin ME-9-58, Metex Mist Eliminators, Metal Textile Corp., 1958, Roselle, N.J.
28. York, Otto H., "Performance of Wire Mesh Demisters", *Chem. Eng. Prog*., August 1954.
29. *Fleximesh Design Manual*, Koch Engineering Co., Inc.
30. Carpenter, C. L. and D. F. Othmer, "Entrainment Removal By a Wire Mesh Separator", *A.I.Ch.E. Journal*, Vol. I, 1955, p. 549.
31. Roberts, E.J., P. Stavenger, J. P. Bowersox, A. K Walton and M. Mehta, "Solids Concentration", *Chem. Eng*., V. 77, No. 14, June 29, 1970.
32. Otto York Co., Bulletin, 55 B, Parsippany, N.J.
33. McNulty, K. J., J. P. Monat and O. V. Hansen, "Performance of Commercial Chevron Mist Eliminators", *Chem. Eng. Prog*., V. 83, No. 5, 1987, p. 48.
34. Watkins, R. N., "Sizing Separators and Accumulators", Hydrocarbon Processing, V. 46, No. 11, 1967, p 253.
35. Branan, C. R., "Pocket Guide to Chemical Engineering", Gulf Publishing Co. Houston, Texas, 1999.
36. Barton, R. L., "Sizing Liquid-liquid Phase Separators Empirically", Chemical Engineering July, 8, 1974.
37. Gerunda, A., "How to Size Liquid-Vapor Separators", Chem. Eng., May 4, 1981, p. 81.
38. Kelsall, M. A., "A Study of the Motion of Solid Particles in a Hydraulic Cyclone", Trans. IChemE., 30, pp. 87–108, 1952.

39. Coker, A. K. and G. V. Jeffreys, "Flow Patterns and Residence Time Distribution In Nozzle Type Reactors", IChemE., Symp. Ser., Fluid Mixing III, 108, p. 105, 1987.
40. Coker, A. K., "A Study of Fast Reactions in Nozzle Type Reactors", Ph.D. Thesis, Aston University, U.K. 1985.
41. Coker, A. K., G.V., Jeffreys and C. J. Mumford, "Mechanisms of Nozzle Type Reactor Operation With Gas Removal", Scientific Res. 49, pp. 67–81, 1992.
42. Koch, W.H. and W. Licht, "New Design Approach Boasts Cyclone Efficiency", *Chem. Eng.*, Nov. 7, 1977, p. 80.
43. Contantinescu, S., "Sizing Gas Cyclones", *Chem. Eng.*, Feb. 20, 1984, p. 227.
44. Doerschlag, C. and G. Miczek, "How to Choose a Cyclone Dust Collector", *Chem. Eng.*, V. 84, No. 4, 1977, p. 64.
45. Stairmand, C. J., "Design and Performance of Cyclone Separators", Trans. Inst. Vol. 29, 1951, p. 356.
46. Strauss, N. Chapter 6, Industrial Gas Cleaning, Pergamon Press, N.Y., 1966.
47. Van Dongen, J. R.J. and A.J. terLinden, "The Application of Gas/Liquid Cyclones in Oil Refining", Paper 56-A-61, *Amer. Soc. Mech. Engrs.*, Presented at. Petroleum Div., Nov. 1956.
48. Zenz, F. A., "Conveyability of Materials of Mixed Particle Size", Ind & Eng. Chem. Fund., 3, p. 65, 1964.
49. Kalen, B. and F. A. Zenz, "Theoretical Empirical Approach to Saltation Velocity in Cyclone Design", AIChE Symp. Series, 70, 137, p. 388, 1974.
50. Walas, S. M., "Chemical Process Equipment-Selection and Design", Butterworths Series, 1988.
51. American Petroleum Institute, Division of Refining, "Cyclone Dust Collectors", Engineering Report, New York, NY.
52. Lieberman, N. P., "Troubleshooting Process Operations", 2nd Ed., PennWell Books, Oklahoma, 1985.
53. Shepherd, C. B. and C. E. Lapple, "Flow Pattern and Pressure Drop in Cyclone Dust Collectors", *Ind. Eng. Chem*. 31, 972, 1939.
54. ter Linden, A. J., Untersuchungen an Zyklonabscheidern, Tonindustrie-Zeitung und Keramische Rundschau, Sonderruckaus T1Z = Zb1. 77, (1953) Heft 34, Seite 49 bis 55.
55. ter Linden, A.J., Der Zyklon alsTropfenabscheider,ChemieIng-Techn.January25 (1953)/Nr. 6.
56. Waliulla, S., "Do-It-Yourself Vortex Breakers", *Chem. Eng.*, V. 95, No. 7. p 108.
57. Pollak, A. and L. T. Work, "The Separation of Liquid from Vapor, Using Cyclones", *Amer. Soc. Mech. Engrs*. 64, 1942, p. 31.
58. Zanker, A., "Hydrocyclones, Dimensions and Performance", *Chem. Eng.*, V. 84, No. 70, 1977, p. 122.
59. Reid, K.J. and Voller, V.R., "Reconciling Hydrocyclone Particle-Size Data", *Chem. Eng.*, V. 90, No. 13, 1983, p. 43.
60. Toy, D. A. and F. M. Bonady, "Guide to Wet Scrubbers", *Chemical Processing*, Oct. 1983, p. 47.
61. Beddow, J. K., "Dry Separation Techniques", *Chem. Eng.*, V. 88, Aug. 10, 1981, p. 69.
62. Boothroyd, R. G., *Flowing Gas-Solids Suspension*, Chapman and Hall, 1971.
63. Kraus, M.N., "Baghouses: Separating and Collecting Industrial Dusts", *Chem. Eng.*, V. 86, No. 8, 1979, p. 94.
64. Kraus, M.N., "Baghouse: Selecting, Specifying and Testing Industrial Dust Collectors", *Chem. Eng.*, V. 86, April 23, 1979, p. 133.
65. Bergmann, L., "Baghouse Filter Fabrics", *Chem. Eng.*, V. 88, No. 21, 1981, p. 177.
66. Parkinson, G., "A Hot and Dirty Future for Baghouses", *Chem. Eng.*, V. 96, No. 4, 1989, p 30.
67. Lawton, D. J. and T. Constantino, "What Makes Cartridge Filters Perform Effectively", *Chemical Engineering Progress*, V. 83, No. 11, 1987, p. 20.
68. Michaels, S. L., "Crossflow Microfilters: The Ins and Outs", *Chem. Eng.*, V. 96, No.1, 1989, p. 84.
69. Shucosky, A. C., "Select The Right Cartridge Filter", *Chem. Eng.*, V. 95, No. 1, 1988, p. 72.
70. Sargent, G. D., "Gas/Solids Separations", *Chem. Eng./Deskbook*, V. 78, No. 4, 1971, p. 11.
71. Sickles, R W., "Electrostatic Precipitators", *Chem. Eng./Deskbook*, V. 75, No. 22, 1968.
72. Bump, R. L., "Electrostatic Precipitators in Industry", *Chem. Eng.*, V. 84, No. 12, 1977, p. 129.
73. Jaasund, S.A., "Electrostatic Precipitators: Better Wet Than Dry", *Chem. Eng.*, V. 94, No. 17, 1987, p. 158.
74. Van Wassen, R. H., *Electrostatic Precipitators for Air Pollution Control*, Wheelabrator Air Pollution Control, Pittsburgh, Pa.
75. American Petroleum Institute, 12J, Specification for Oil and Gas Separators, 8th ed., October 2008.
76. Bothamley, M., Gas–Liquid Separators—Quantifying Separation Performance Part 1., SPE Oil and Gas Facilities, Aug (22–29), 2013.
77. Srikumar Koykkal, Chemical Process Technology and Simulation, PHI Learning Private Ltd., Delhi 110092, 2013.
78. Campbell, John, M., Robert A. Hubbard, and Robert A. Maddox, Gas Conditioning and Processing, Campbell Petroleum Series.
79. Ken E. Arnold, and Maurice Stewart, Surface Production Operations, Vol. 1, Elsevier, 1999.
80. PF-49, Troubleshooting Oil and Gas Processing Facilities, Bothamley, M., Petro Skills, LLC. 2014.
81. Ahmed, T., Makwashi, N., and M. Hameed, A Review of Gravity Three-Phase Separators, Jour. Of Emerging Trends in Engineering and Applied Sciences, p. 143, 2017.
82. Gas Processing Suppliers Association (GPSA), Engineering Data Book, 12th Ed. 2008.

83. Campbell, J. M., Gas Conditioning and Processing, Vol. 2: The Equipment Modules, 9th edition, 2nd Printing Editors, Hubbard, R., and Snow-McGregor, K., Campbell Petroleum Series, Norman, Oklahoma, 2014.
84. Arbuthnott, Ross, Blanc, K., Haig, M., Lynch, R., Robertson, M., and S. Van der Hem., How to Size and Rate Horizontal Three-Phase Separators, The Chemical Engineer, pp. 48–52, February 2017.
85. Srikumar, K., Oil and Gas Pipeline Developments and Optimisation, Chemical Industry Digest, December 2007.
86. Muravyov, I., Andriasov, R., and V. Poloskov, Development and Exploitation of Oil and Gas Fields, Peace Publishers, Moscow.

19

Distillation

19.1 Distillation Process Performance

An important unit operation in the chemical process industries (CPI) is to separate a mixture into its components. This is termed diffusional or mass transfer. A typical chemical plant as illustrated in the schematic flow sheet of Figure 19.1 will consist of both reaction and separating units. The raw materials are first purified in a separating unit and then fed to the reactor where they undergo chemical transformation into the main reaction products and possibly into by-products that take place under specified conditions. A factor representing the efficiency with which raw materials are converted to valuable product is the selectivity. This is given by moles of primary product produced divided by the moles of limiting reactant consumed. It can vary between 0 to 1, depending on the stoichiometry, the molar feed ratio of reactants, reactor temperature(s), reactor configuration, the catalyst if required and so on. Any unreacted feed that remains is separated from the reaction products and recycled back to the reactor. The products are further separated and purified before marketed or used in subsequent processes. If the products contain "non-condensable" components such as methane, hydrogen, argon, these must be separated by flash separation or similar process. Before flash separation, the process stream is usually cooled and depressurized. Afterward, it may be fed to another separator to remove and purify useful components. Any remaining raw materials are recycled to the chemical process. Alternatively, the vapor stream may be used as fuel or when valuable raw materials are present in the vapor, the stream may be purged and the gas is recycled to the chemical process. The arrangement of Figure 19.1 is typical of many chemical processes and is illustrated in Hydrocarbon processing magazine [1] and many text books [2–5].

The main techniques used for separations are distillation, absorption, liquid–liquid extraction, drying, leaching, crystallization and gas adsorption. These processes account for 40% to 70% of both the capital and operating costs in the CPI since they significantly affect energy consumption, product costs and manufacturing profits. About 43% of the energy consumed by the U.S. chemical process industries is used for separation processes [6].

The optimum method for a given process depends on the physical state (e.g., liquid, solid or gaseous) of the components such factors as the achievable degree of separation, the required volatility and toxicity of the components and purity criteria. In general, the choice of the key design parameters will need to balance costs incurred due to the raw materials, the reactor system and the separation processes. For example, in some cases, the selectivity increases as the conversion of the limiting reactant decreases. Increased selectivity means that less raw material is converted to by-product, thereby saving on raw material costs. However, as conversion decreases, more unreacted raw material must be separated from the product stream before being recycled to the reactor. The related costs increase quite

---

A. Kayode Coker. Petroleum Refining Design and Applications Handbook Volume 3, (249–702) © 2022 Scrivener Publishing LLC

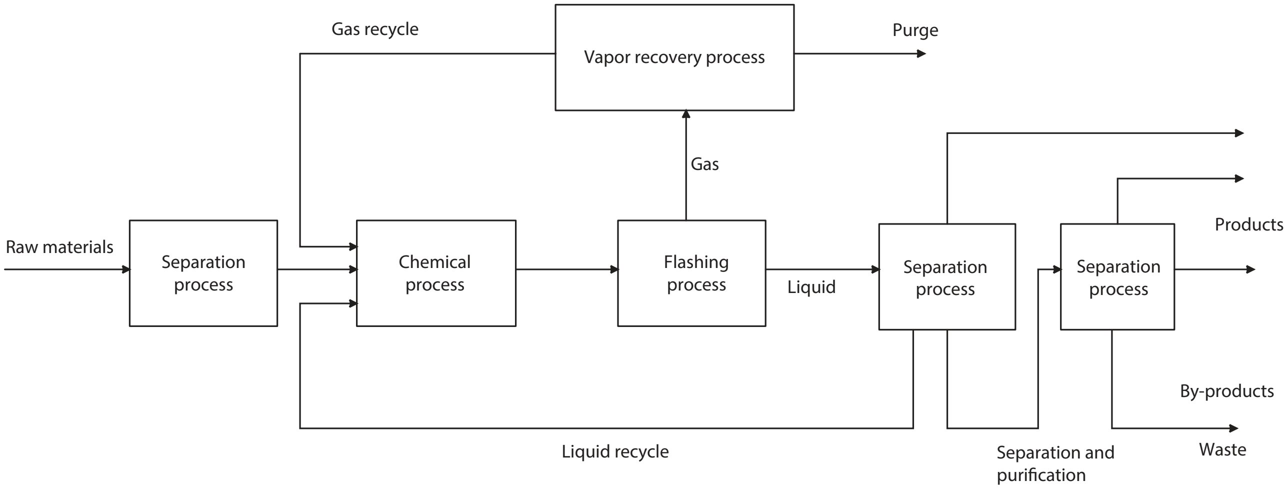


Figure 19.1 Schematic flow diagram of a chemical process.

steeply as conversion decreases. Conversely, reactor volume and its cost increase very sharply as conversion approaches unity. An optimum value therefore must balance the raw material costs, reactor costs and the separation costs. The chosen value will determine the stream temperature, pressure and composition as well as the size of both reactor and the separation system [7].

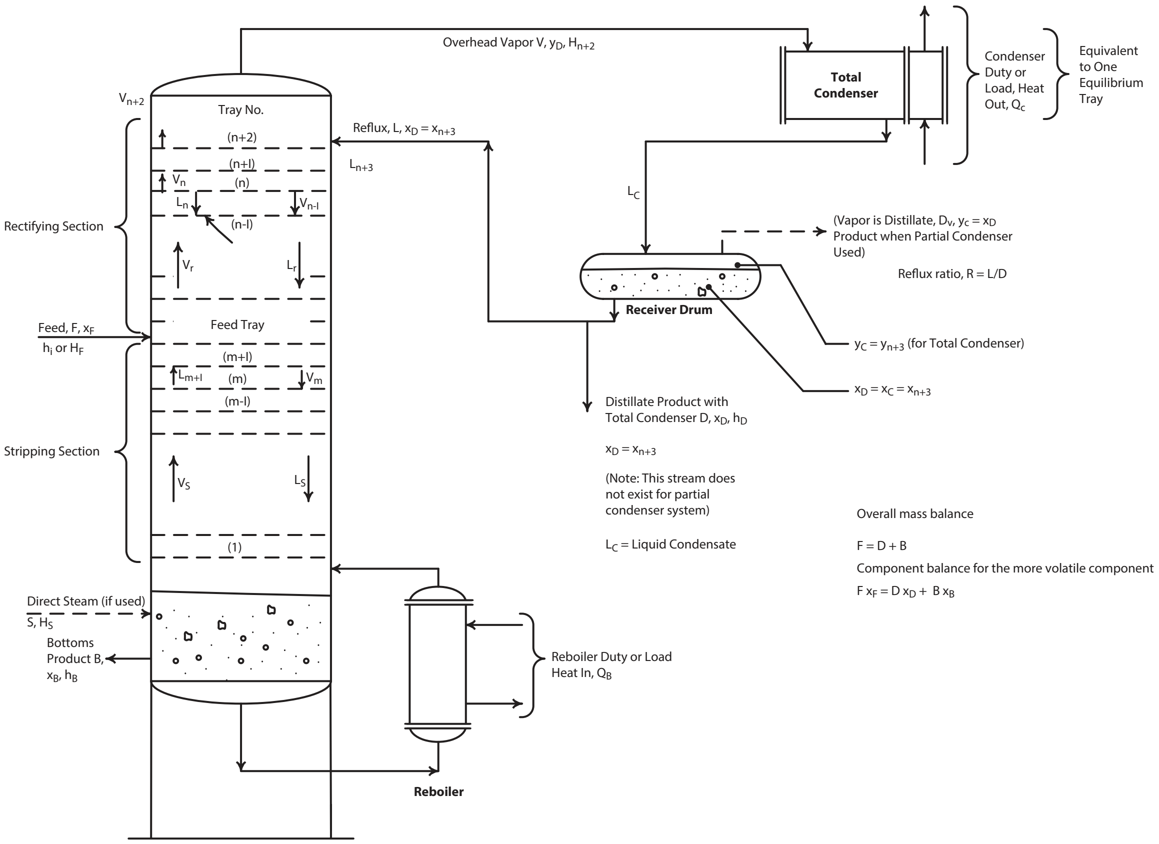


Figure 19.2 Schematic distillation tower/column arrangement with total condenser.

Although the art and science of distillation have been practiced for many years, studies continue to determine the best design procedures for multicomponent, azeotropic, batch, multidraw, multifeed, and other designing. Some shortcut procedures are adequate for complex situations of many systems, but have limitations in others.

The methods outlined in this chapter are considered adequate for the stated conditions, but there may be exceptions to these generalizations. The process engineer often "double checks" his or her results by using a second method to verify the "ball-park" results, or to check the findings of shortcut methods that lack fine detail.

Current design techniques using computer programs and simulation software allow excellent prediction of performance for complicated multicomponent systems such as azeotropic or high hydrocarbon content process streams as well as those which specify extremely high purity of one or more product streams. Of course, the more straightforward systems can be predicted with excellent accuracy. Such techniques allow examination of an array of variables; invaluable in selecting optimum or at least preferred modes or conditions of operation.

The high manufacture costs associated with an installation of this equipment warrant the best design and checking techniques. The process diagram of Figure 19.2 will be used as reference for the systems and methods presented in this chapter; Figure 19.2a shows a photograph of fractionating columns in a chemical plant.

The right size of equipment should be chosen for the process. In addition, a choice between a stagewise or continuous, i.e., tray or packed process needs to be made. Trays act as individual stages and produce stepwise changes in concentration, whereas packed towers produce continuous differential changes in concentration.

This chapter reviews the vapor–liquid equilibria of some representative mixtures, conditions of bubble point of liquid mixtures and dew point of gaseous mixtures, isothermal equilibrium flash calculations, the design of distillation towers with bubble caps, valve trays, packed tower design (see chapter 20), Smoker's equation for estimating the

Figure 19.2a A photograph of fractionating columns showing De-methanizer, De-ethanizer, De-propanizer, and De-butanizer units (courtesy of Sulzer Chemtech Ltd.).

number of plates in a binary mixture, petroleum fractionation, reactive distillation, the computation of multicomponent recovery and minimum trays in distillation columns troubleshooting, distillation sequencing, and heat integration of distillation columns. Computer, Excel spreadsheet programs and Honeywell UniSim simulation software are used to illustrate various examples in this volume.

19.2 Equilibrium Basic Considerations

Distillation design is based on the theoretical premise that heat and mass transfer from stage to stage (theoretical) are in equilibrium [8–12]. Actual columns with actual trays are designed by establishing column tray efficiencies, and applying these to the theoretical trays or stages that have been determined by the calculation methods to be presented in later sections.

Dechman [13] illustrates a modification to the usual McCabe–Thiele diagram that allows for unequal molal overflow. Ryan [14] provides an alternative McCabe–Thiele diagram for relative volatilities below 1.25. The procedure of generating the revised McCabe–Thiele diagram is provided later in this volume.

Distillation, extractive distillation, liquid–liquid extraction and absorption are all techniques used to separate binary and multicomponent mixtures of liquids and vapors. Reference [15] examines approaches to determine optimum process sequences for separating components from a mixture, primarily by distillation. The most important assumption is that vapor and liquid are in equilibrium to each theoretical stage. Heat and mass transfer are complete and at equilibrium, meaning that the temperature of both phases is the same. The composition in each phase is uniform.

Vapor–liquid equilibrium data is therefore essential to adequately perform distillation calculations and so to correlate composition, temperature and system pressure. There are two types of systems: ideal and non-ideal. These terms apply to the simpler binary or two component systems as well as to the often more complex multicomponent systems.

Figure 19.3 shows a photograph of a typical vapor–liquid equilibrium apparatus with controls. Figures 19.4a and 19.4b show a photograph of a pilot plant distillation column and its schematics, respectively; Figures 19.4c and 19.4d show a photograph of the column plates and its schematics with the direction of vapor and liquid flows.

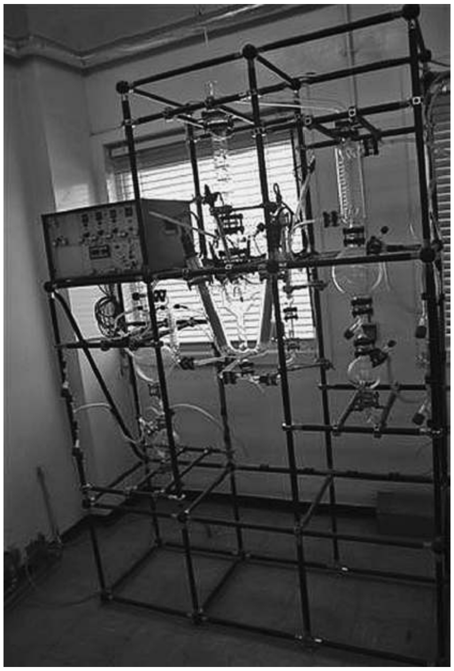
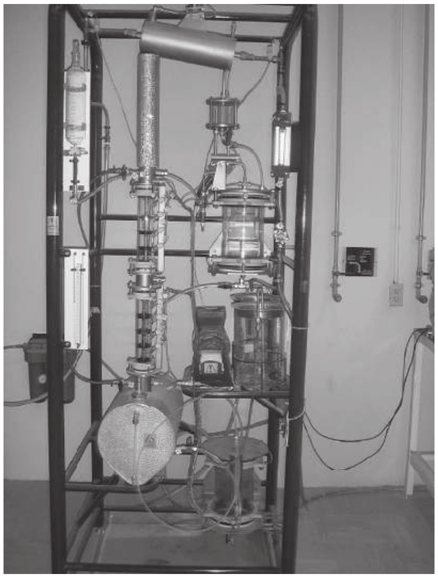


Figure 19.3 A photograph of vapor-liquid equilibrium apparatus (courtesy of Saudi Aramco Shell Refinery Co. (SASREF)).



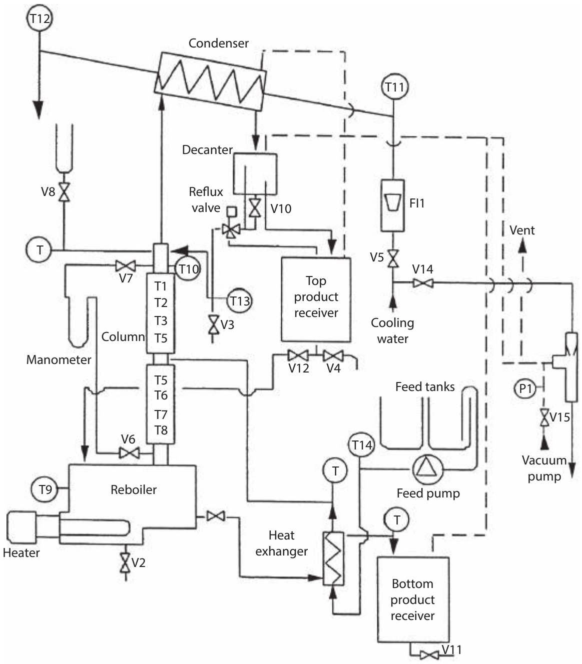
(**a**)

Figure 19.4a Photograph of a pilot plant distillation column (courtesy of Armfield U.K.).

Figure 19.5 illustrates a typical volatility vapor-liquid equilibrium curve for a particular component in a distillation separation. Usually for the more volatile component (MVC) of the mixture is the one of interest.

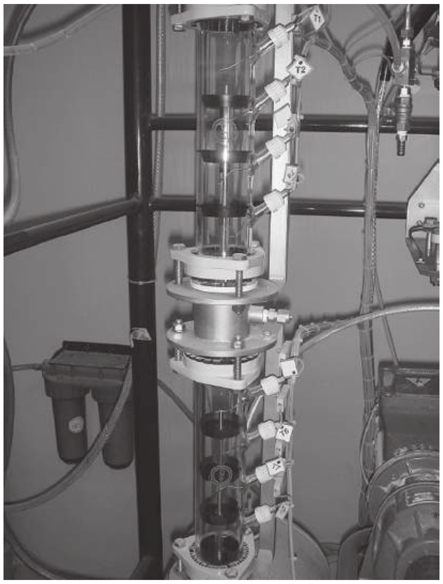
19.3 Vapor–Liquid Equilibria

Vapor-liquid equilibrium (VLE) data form the basis for evaluation of distillation operations; the most useful form of this data are the boiling point diagrams. An ideal liquid or gas solution is one for which the activity coefficient is



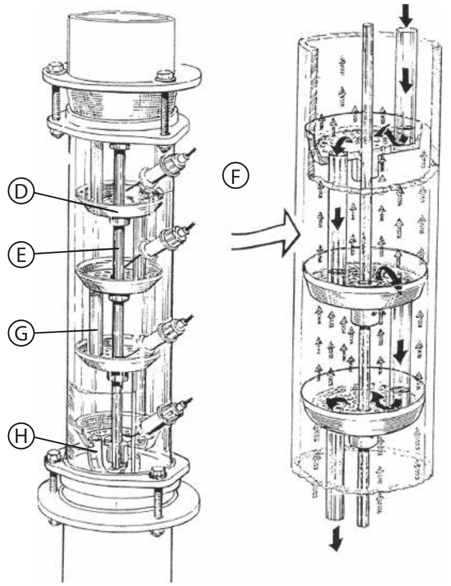
(**b**)

Figure 19.4b Schematics of a pilot plant distillation column (courtesy of Armfield, U.K.).



(**c**)

Figure 19.4c Photograph of the column plates (courtesy of Armfield, U.K.).



(**d**)

Figure 19.4d Schematics of a pilot plant distillation column plates showing liquid and vapor movements (courtesy of Armfield, U.K.). D = a plate; E = a support rod; F = a weir; liquid movement ⬇ and vapor movement ⇧, G = a downcomer; H = U-tube.

unity. There are few liquid systems that behave ideally at any pressure, and the assumption of ideal liquid behavior is generally correct only for members of homologous series that have similar molecular weights. At low pressure, the vapor phase of a mixture approaches ideal behavior and follows the ideal gas law. For these conditions, the fugacity is equated with the partial pressure and the standard state fugacity of the liquid becomes equal to the vapor pressure.

Figure 19.6a is a phase diagram for a mixture of methanol and water at a constant pressure of one atmosphere (101.3 kPa). The boiling points of pure methanol, the more volatile component (MVC) and pure water are represented by temperatures 64.5°C and 100°C, respectively. The lower curve represents the locus of the bubble point temperatures

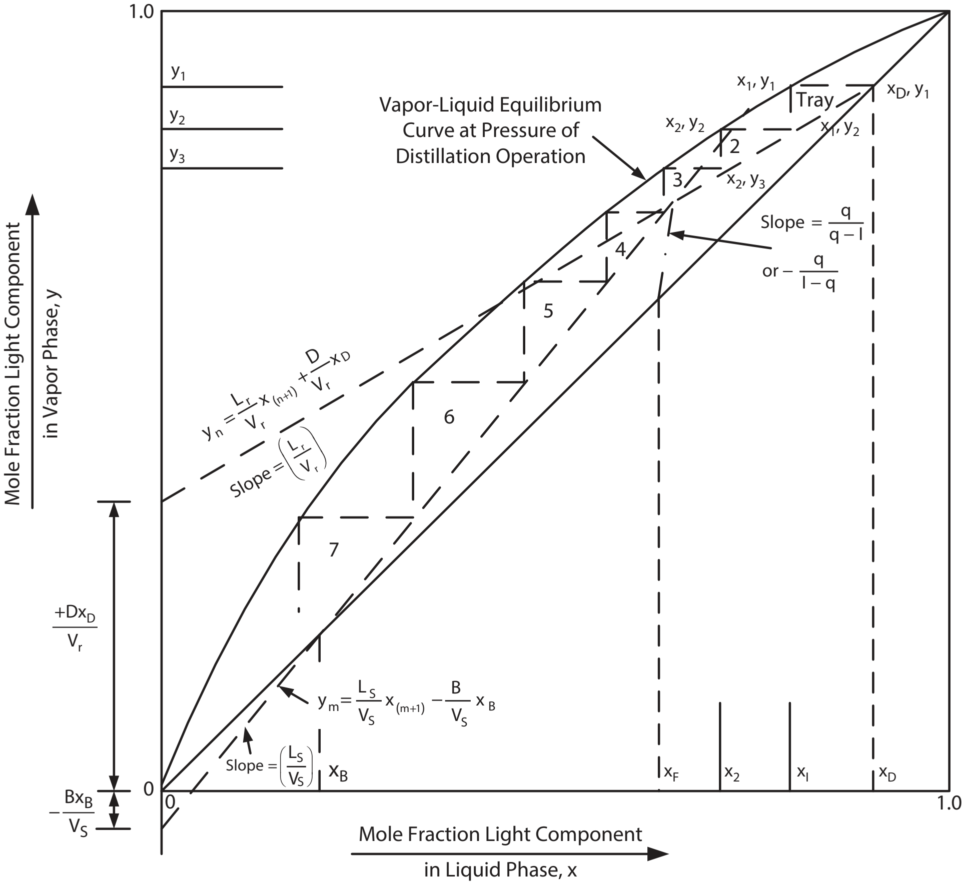


Figure 19.5 Continuous fractionation of binary mixtures: McCabe-Thiele diagram with total condenser.

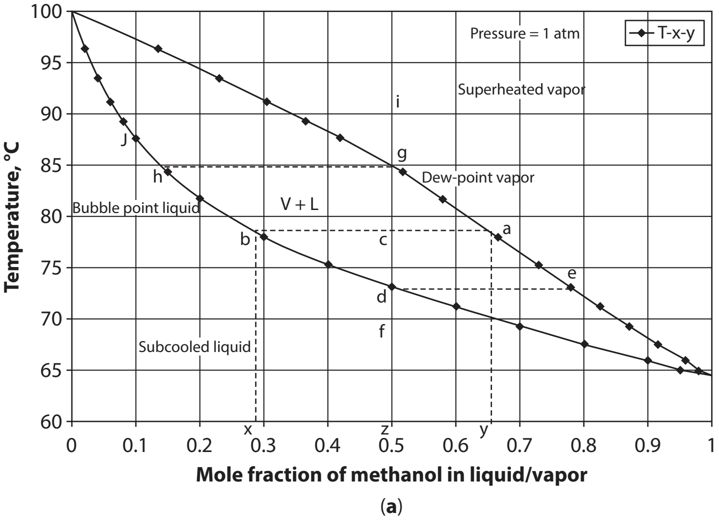


Figure 19.6a Phase diagram for the mixture of methanol and water at 1 atm.

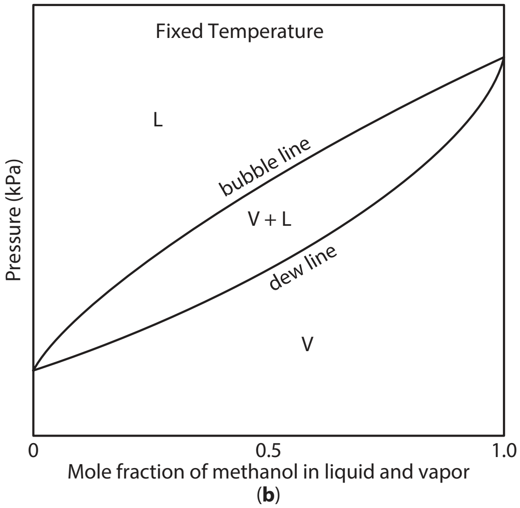


Figure 19.6b Illustration of P-x-y diagram.

for all mixture of methanol and water. This curve is termed the saturated liquid (bubble point liquid); states of subcooled liquid lie below it.

The upper curve represents the locus of the dew-point temperatures; states of superheated vapor lie above it. This curve is termed the saturated vapor (dew-point vapor). Both of these curves originate at the boiling temperature of one constituent (e.g., 100°C) and terminate at the boiling temperature of the other constituent (64.5°C). The two-phase region lies between these curves.

Assume that a mixture of methanol-water of composition at point *z* (50 mole %) exists at a temperature of 70°C (represented by point ***f*** on Figure 19.6a). As this mixture is heated, it will remain all liquid until point ***d*** (74°C) is reached, at which the first bubble of vapor is formed. This is the bubble-point temperature for a 50 mole % mixture at the system pressure of 1 atmosphere. The vapor phase composition at the bubble point temperature is 78 mole % of methanol (point ***e***). If heating of the mixture is continued, and the vapor generated is maintained in equilibrium with the liquid, the mixture reaches a point ***c*** on the diagram, at which both vapor and liquid exist. The overall composition of the total mixture remains ***z***, but the entire liquid portion will have a composition ***x*** (28 mole %), and the vapor portion will have a composition *y* (66 mole %). However, once a small amount of liquid has been vaporized, the remainder no longer contains 50 mole % methanol; it contains less, as the vapor produced has a higher concentration of this component. Consequently, the temperature of the system steadily rises as more and more liquid is vaporized, and the compositions of both phases change continuously during the process. The vapor and liquid phase compositions change as indicated by paths ***bh*** and ***ag***, until the dew point is reached at point ***g***, where the last bubble of liquid disappears.

Correspondingly, consider an experiment started with 50 mole % mixture at 90°C (point ***i***), where the mixture is 100% vapor. As the mixture temperature is lowered, the dew-point temperature is encountered at 85°C at system pressure (point ***g***), and the first drop of liquid is formed which is about 15 mole % (point ***h***). As the temperature of the system steadily decreases, more and more vapor is condensed, and the compositions of both phases change continuously as indicated by paths ***ga*** and ***hb***, until the bubble point is reached at point ***d***, where the last bubbles of vapor disappear.

The bubble and dew temperatures are composition-dependent. For example, the bubble temperature of a 28 mole % mixture is 77°C (point ***b***) and the bubble temperature of a 10 mole % is 87°C (point ***j***). A similar discussion to that

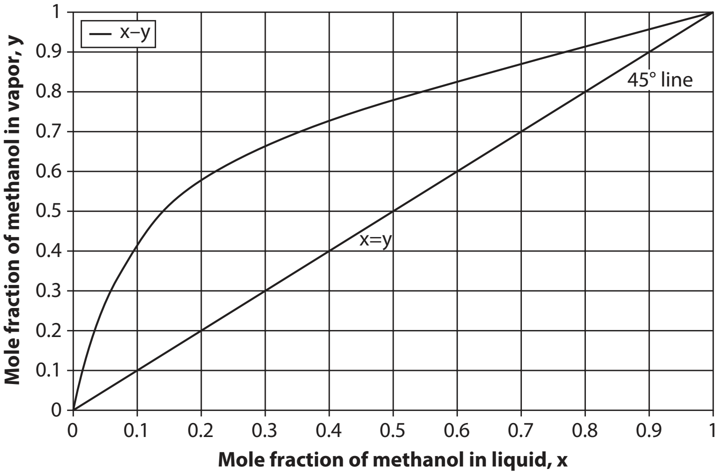


Figure 19.7 VLE diagram for the mixture of methanol and water at 1 atm.

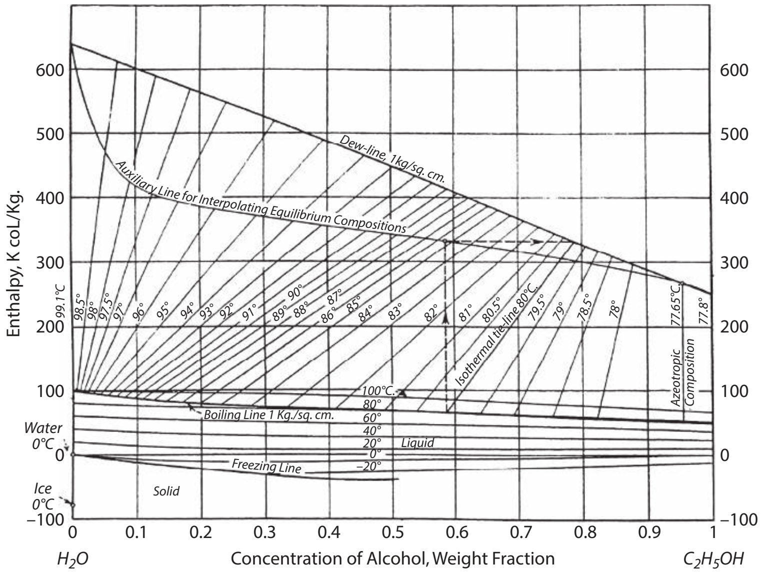


Figure 19.8 Enthalpy-composition diagram for ethanol-water at a pressure of 1 kg/cm^2 (source: Bosnjakovic, Technishche Thermodynamik, T. Steinkopff, Leipzig, 1935).

given above could be presented for the dew temperatures of these systems. The bubble and dew-point discussions could also be presented on the pressure diagram at constant temperature as shown in Figure 19.6b; however the references are to the bubble and dew pressures. The horizontal dotted lines connecting coexisting compositions are tie lines, and the relative amounts of the vapor and liquid present can be determined by the material balances.

In Figure 19.6a, any point above the saturated vapor curve represents an all vapor mixture (superheated). Correspondingly, all points below the saturated liquid curve represent an all liquid condition (subcooled), and points between the two curves correspond to a liquid–vapor mixture. Where the boiling points diagram is unavailable for a particular system, equilibrium relationships are obtained from vapor pressures data using Raoult's law and Dalton's law, where ideal behavior may be assumed. Recently, simulation software such as Chemcad, Aspen plus, HYSYS, and UniSim (Honeywell) have incorporated thermodynamic properties in their workbench and thus allowed phase diagrams of mixtures (T-x-y, x-y, etc.) to be readily generated. Various VLE diagrams of mixtures have also been constructed using equilibrium data from Perry's Chemical Engineers' handbook [16] and chemical engineering

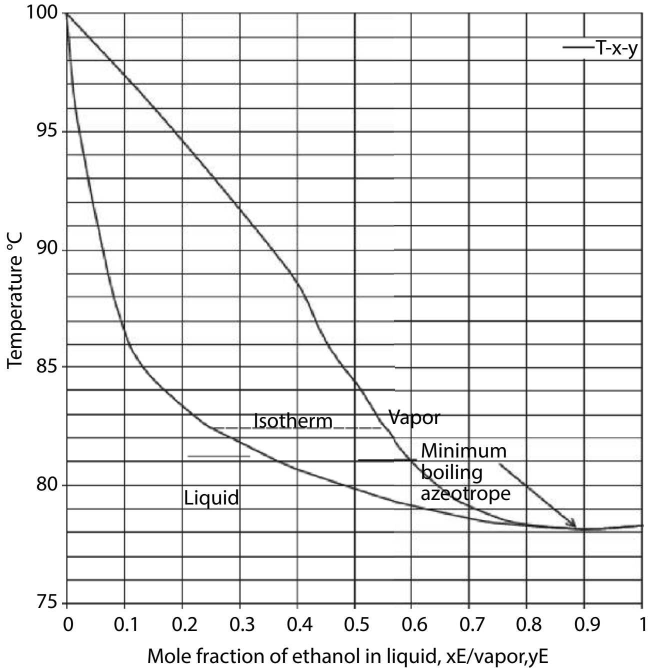


Figure 19.8a Temperature composition diagram for ethanol and water mixture at 1.013 kPa.

thermodynamic texts [17–20]. These phase diagrams can be accessed in this volume, and are available from Wiley–Scrivener companion web sites. Figure 19.7 shows the VLE (x-y) diagram for the mixture of methanol and water at one atmosphere, generated from the T-x-y data of Figure 19.6a. Figure 19.8 shows the enthalpy composition diagram for ethanol–water mixture; Figure 19.8a shows the temperature composition diagram of ethanol–water with a minimum boiling azeotrope at 78.2°C, and Figure 19.7 forms the basis of McCabe–Thiele diagram for calculation of continuous fractionation of binary mixtures; this will be presented in the graphical design of a distillation column.

The distillation process depends on the differences between the compositions of liquid and vapor phases of a mixture at equilibrium. This implies that the temperatures and pressures of the phases must be the same, and no composition change occurs with time. Equilibrium can be achieved after a long period of thorough mixing and contact between the two phases. An ideal gas is by definition one that follows the ideal gas laws:

$$PV = nRT \tag{19.1}$$

where

P = absolute pressure
V = total volume
n = number of moles of gas
R = ideal gas constant
T = absolute temperature

The conditions under which a given component or mixture approaches ideal behavior depend on its critical temperature and critical pressure. Other applicable principles of gas behavior are Dalton's law of additive pressures and Amagat's law of additive volumes. These are:

$$\pi = p_A + p_B + \ldots \tag{19.2}$$

and

$$V_T = V_A + V_B + \dots \tag{19.3}$$

where

π = total system pressure
V_A = volume of component A
V_B = volume of component B
V_T = total volume of system
p_A = partial pressure of component A
p_B = partial pressure of component B
A, B = components

These laws are only applicable when conditions are such that each component and the mixture obey the ideal gas law. The equilibrium between two phases can be related to the equality of the chemical potential and defined in terms of the Gibbs free energy as

$$\mu_i = \left[\frac{\partial G}{\partial \mu_i}\right]_{T,P,n_i} = \bar{G}_i \tag{19.4}$$

where $\bar{G}_i$ is the partial molar Gibbs free energy. From the fundamental thermodynamic relation

$$dG = VdP - SdT \tag{19.5}$$

The effect of pressure is mostly observed at constant temperature as defined by

$$dG = VdP \text{ (constant T)} \tag{19.6}$$

where G is the molar Gibbs free energy and V is the molar volume. For a pure fluid of component i and substituting Equation 19.1 into Equation 19.6 yields,

$$dG = RT\frac{dP}{P} = RT\,d(\ln P) \tag{19.7}$$

The free energy from Equation 19.7 is restricted to ideal gases, but can be modified to cover real fluids by introducing a new function defined as the fugacity. Equation 19.7 becomes

$$dG_i = RT\ d(\ln f_i) \tag{19.8}$$

where f_i is the fugacity of pure i with units of pressure. The fugacity becomes equal to the pressure as the pressure approaches zero.

Thus

$$\lim \frac{f_i}{P} = 1.0 \qquad P \to 0 \tag{19.9}$$

For a single component in a mixture at constant temperature, the fugacity is defined by

$$d\bar{G}_i = RT\, d(\ln \bar{f}_i) \tag{19.10}$$

and

$$\lim_{P \to 0} \frac{f_i}{y_i P} = 1.0 \tag{19.11}$$

If a liquid mixture at temperature T and pressure P is in equilibrium with a vapor mixture at the same temperature and pressure, therefore at an equilibrium condition, with the Gibbs free energy minimized, the thermodynamic criterion will be

$$\hat{f}_i^V = \hat{f}_i^L \; (i = 1, 2\ldots\ldots\ldots N) \tag{19.12}$$

The vapor fugacity can be regarded as a corrected partial pressure, and given by

$$\hat{f}_i^V = \phi_i^V (P y_i) \tag{19.13}$$

where

$\hat{f}_i^V$ = fugacity of component i in the vapor phase
ϕ_i^V = the vapor fugacity coefficient of species i, based on P, T, y_1, y_2.
N = number of chemical species
y_i = mole fraction of species i in the vapor phase.
P = system total pressure.

The fugacity of a component in a mixture depends on the temperature, pressure and composition. The fugacity $\hat{f}_i^L$, of any component i, in the liquid phase is related to the composition (mole fraction) of the component in that phase by:

$$\hat{f}_i^L = \gamma_i x_i f_i^{oL} \tag{19.14}$$

Similarly, the liquid fugacity can be regarded as a corrected vapor pressure by

$$\hat{f}_i^L = \phi_i^L \gamma_i \psi_i \left(x_i p_i^o\right) \tag{19.15}$$

where

ϕ_i^L = the liquid fugacity coefficient of species i based on T and p_i^o for pure liquids.
ψ_i = the Poynting correction factor for component i accounting for the effect of pressure on liquid phase fugacity
γ_i = the liquid phase activity coefficient of species i
x_i = mole fraction of species i in the liquid phase
$\hat{f}_i^L$ = fugacity of species i in the liquid phase
f_i^{oL} = the fugacity of species i in the reference (or standard state).

The fugacity of pure liquid i at temperature T and pressure P is given by

$$f_i^{oL} = P_{vp,i}(T)\phi_i^s(T)\exp\left[\int_{P_{vp,i}}^{P} \frac{V_i^L(T,P)}{RT}\,dP\right] \tag{19.16}$$

where $P_{vp,i}$ = the vapor pressure of component i
ϕ_i^s = the vapor phase fugacity coefficient of pure saturated vapor of component i

The exponential term in Equation 19.16 is a correction factor for the effect of pressure on liquid-phase fugacity and is termed the **Poynting correction factor**. In Equation 19.16, V_i^L can be replaced by the partial molar volume of component i in the liquid solution for greater accuracy. For low to moderate pressure, V_i^L is assumed as the saturated liquid molar volume at the specified temperature. Equation 19.16 is simplified to give

$$f_i^{oL} = P_{vp,i}\,\phi_i^s \exp\left[\frac{V_i^{Ls}(P - P_{vp,i})}{RT}\right] \tag{19.17}$$

Equation 19.17 is used to calculate the reference state fugacity of liquids. Any equation of state can be used to determine ϕ_i^s. For low to moderate pressures, the virial equation is the simplest to use. The fugacities of pure gases and gas mixtures are required for estimation of many thermodynamic properties, such as entropy, enthalpy and fluid phase equilibria. For pure gases, the fugacity is

$$\ln f = \frac{1}{RT}\int_0^P V\,dP \tag{19.18}$$

The fugacity can be calculated from Equation 19.18 once the P-V-T behavior of the fluid is known from an equation of state. The fugacity coefficient is the ratio of the fugacity of a substance to its pressure. For a pure substance,

$$\phi = \frac{f}{P} \tag{19.19}$$

For the vapor phase, the composition is generally expressed by the mole fraction, y_i

$$\phi_i = \frac{f_i^V}{y_i P} \tag{19.20}$$

For a mixture of ideal gases, $\phi_i = 1$. The fugacity coefficient, ϕ_i, depends on temperature and pressure. The fugacity coefficient is normalized, such that as $P \to 0$, $\phi_i \to 1$ for all components. At a low pressure, ϕ_i is assumed to be unity. When the relationship between volume and pressure is known, either from direct measurement or from an empirical equation of state, changes in fugacity may be found by integration. Thus

$$\ln\frac{f_2}{f_1} = \frac{1}{RT}\int_{P_1}^{P_2} V\,dP \tag{19.21}$$

Since $\phi = 1$ when $P = 0$, absolute values of the fugacity coefficient can be determined. Therefore,

$$\ln\phi = \ln\frac{f}{P} = \frac{1}{RT}\int_0^P\left(V - \frac{RT}{P}\right)dP$$
$$= \int_0^P (Z-1)\frac{dP}{P} \qquad (19.22)$$

Methods for estimating fugacities of liquid and vapor phases are given by Reid *et al.* [21] and by Walas [17] and other thermodynamic texts [18, 19]. These also give detailed derivation of the above equations and their relationships. The various coefficients in these equations are illustrated below [22]:

ϕ_i^V	The vapor fugacity coefficient accounts for the effect of vapor non-ideality on vapor fugacity. It is usually estimated from an equation of state and is based on system temperature, pressure, and vapor mole fraction.
ϕ_i^L	The liquid fugacity coefficient accounts for the effect of vapor non-ideality on liquid fugacity. It is estimated in a similar manner to the vapor fugacity coefficient, but it is based on the system temperature and the pure component vapor pressure.
ψ_i	The Poynting correction factor accounts for the effect of pressure on liquid fugacity. Since ϕ_i^L is evaluated at the vapor pressure of the pure component, ψ_i is used to account for the difference between the pure component vapor pressure and the mixture pressure. This effect is small and can be neglected at low pressures, but is important at high pressures.
γ_i	The liquid activity coefficient corrects the liquid fugacity for the effect of composition. Its value depends on how similar the components are. For two similar components, such as an isobutane–normal butane mixture, the liquid activity coefficient is close to unity. If the components are different, activity coefficients deviate from unit.

19.4 Activity Coefficients

Dissimilar molecules repel each other resulting in activity coefficients (γ_i) greater than 1. So, in a mixture of unlike molecules (e.g., polar water molecules are mixed with nonpolar organic molecules) greater partial pressure is exerted resulting in a positive deviation from ideality. Figures 19.9a and 19.9b show typical plots of activity coefficients γ_1 and γ_2 of the two components of such a mixture.

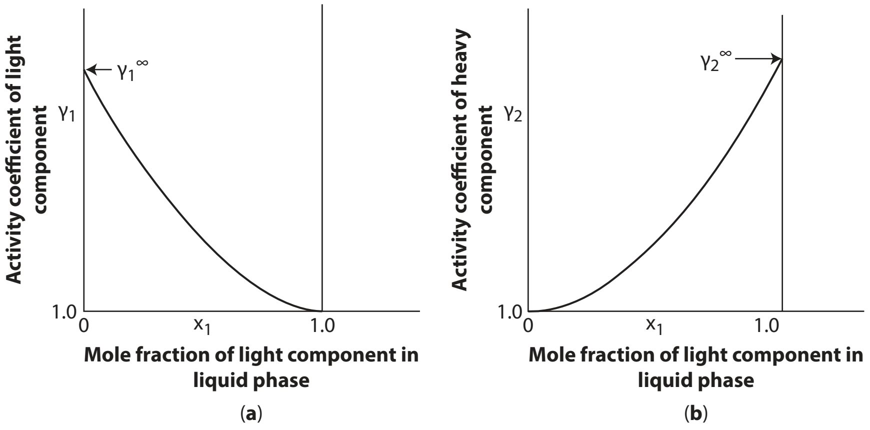


Figure 19.9 (a, b) Composition dependence of activity coefficient showing positive deviation from ideality.

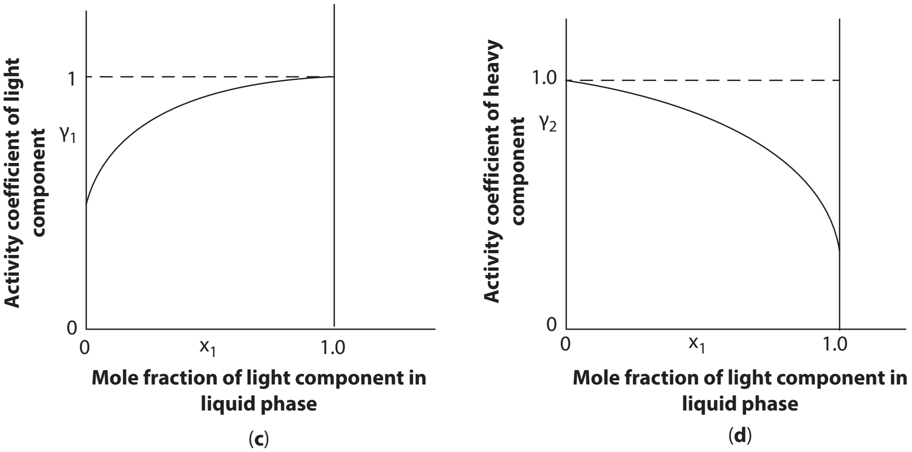


Figure 19.9 (c, d) Composition dependence of activity coefficient showing negative deviation from ideality.

At $x_1 = 0$; the pure heavy component (component 2) has an activity coefficient (γ_2) of unity. The activity coefficient (γ_1) of light component (component 1) will be large and is termed the infinite dilution activity coefficient γ_1^∞.

At $x_1 = 1$, the activity coefficient (γ_1) of light component (component 1) is unity. The activity coefficient of heavy component (component 2) is the infinite dilution activity coefficient γ_2^∞.

In the opposite case, where the molecules of the components attract each other, the activity coefficients will be less than 1 as the molecules will exert lower partial pressure in a mixture than if they were pure. For example, when nitric acid is mixed with water, a lower pressure is exerted and Figures19.9c and 19.9d show a typical activity coefficient dependence on composition for this case of negative deviation from ideality.

19.5 Excess Gibbs Energy—G^E

The excess Gibbs energy $\bar{G}_i^E$ can be expressed by

$$\bar{G}_i^E = RT \ln \frac{\hat{f}_i}{x_i f_i} \tag{19.23}$$

For an ideal solution, the fugacity of each species is proportional to its mole fraction, i.e.:

$$\hat{f}_i = x_i f_i \tag{19.24}$$

where the proportionality constant is the fugacity of pure species i in the same physical state as the solution and at the same temperature, T and pressure, P. Equation 19.24 is known as the ***Lewis/Randall rule*** and it applies to each species in an ideal solution at all conditions of temperature, pressure and composition. The dimensionless ratio $\hat{f}_i / x_i f_i$ is the activity coefficient of species i in solution and is represented by $\boldsymbol{\gamma}_i$. This is expressed by:

$$\gamma_i \equiv \frac{\hat{f}_i}{x_i f_i} \tag{19.25}$$

Substituting Equation 19.25 into Equation 19.23 gives

Table 19.1 VLE data for methyl ethyl ketone (1)/toluene (2) at 50°C.

P/kPa	x_1	y_1	$\hat{f}_1 = y_1 P$	$\hat{f}_2 = y_2 P$	γ_1	γ_2
12.30 (P_2^{sat})	0.0000	0.0000	0.000	12.300		1.000
15.51	0.0895	0.2716	4.212	11.298	1.304	1.099
18.61	0.1981	0.4565	8.496	10.114	1.188	1.026
21.63	0.3193	0.5934	12.835	8.795	1.114	1.050
24.01	0.4232	0.6815	16.363	7.697	1.071	1.078
25.92	0.5119	0.7440	19.284	6.636	1.044	1.105
27.96	0.6096	0.8050	22.508	5.542	1.023	1.135
30.12	0.7135	0.8639	26.021	4.099	1.010	1.163
31.75	0.7934	0.9048	28.727	3.023	1.003	1.189
34.15	0.9102	0.9590	32.750	1.400	0.997	1.268
36.09 (P_1^{sat})	1.0000	1.0000	36.090	0.000	1.000	

(Source: M. Diaz, Pena, A. Crespo Colin and A. Compostizo, J. Chem. Thermodyn, vol. 10, pp 337-341, 1978).

$$\bar{G}_i^E = RT \ln \gamma_i \tag{19.26}$$

These equations establish a thermodynamic relationship between the excess Gibbs energy and activity coefficient. For a binary system, the excess Gibbs energy is expressed by

$$\frac{G^E}{RT} = x_1 \ln \gamma_1 + x_2 \ln \gamma_2 \tag{19.27}$$

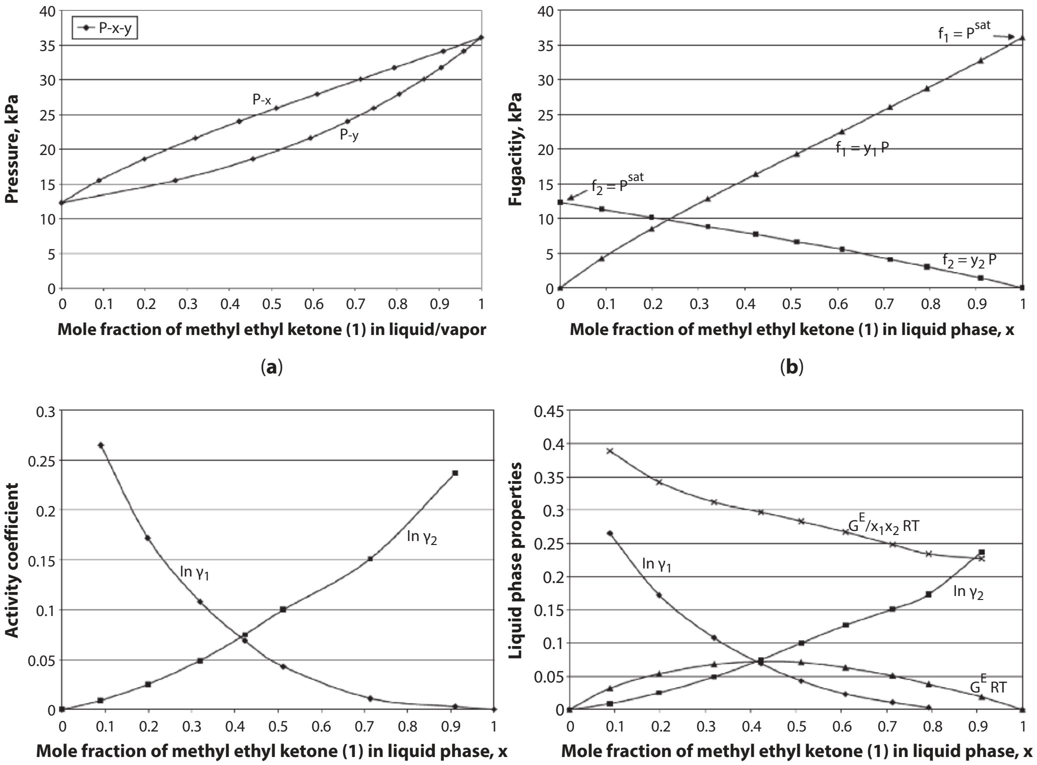
Table 19.1 shows the vapor–liquid equilibrium data for methyl ethyl ketone (1) and toluene (2) mixture at 50°C, and Figures 19.10a, 19.10b, 19.10c, and 19.10d, respectively, show plots of pressure against compositions of methyl ethyl ketone (1) in liquid and vapor phases, fugacity against the mole fraction of methyl ethyl ketone (1) in liquid phase and the liquid phase properties of methyl ethyl ketone (1) and toluene (2) mixture. These plots use the experimental data of Pena *et al.* [23] and developed using Excel spreadsheet (Phase-Equil.xls). The detailed explanation and derivation of GE and activity coefficient γ_i are given in standard chemical engineering thermodynamic texts [18, 19].

19.6 K-Value

The equilibrium constant K can be defined as:

$$K_i = \frac{\text{Mole fraction of component i in vapor phase}}{\text{Mole fraction of component i in liquid phase}} \tag{19.28}$$

where the K-value is a measure of the tendency of component i to vaporize. It is a function of temperature, pressure and composition. At equilibrium, whenever two of these three variables are fixed, the third variable is also fixed. Therefore, the K value can be defined as a function of composition and pressure; composition and temperature or



(**a**) (**b**)

(**c**) (**d**)

Figure 19.10 (a) Phase diagram for methyl ethyl ketone (1) and toluene (2) mixture at 50°C. (b) Fugacities of methyl ethyl ketone (1) and toluene (2) mixture at 50°C. (c) Logarithms of the activity coefficients for methyl ethyl ketone (1) and toluene (2) at 50°C. (d) Liquid phase properties of methyl ethyl ketone (1) and toluene (2) mixture at 50°C.

temperature and pressure. If the K-value is above 1.0, the component tends to concentrate in the vapor phase, correspondingly, below 1.0, the component tends to concentrate in the liquid phase. If the K-value is unity, the component concentration equal in the vapor and liquid phases.

Combining Equations 19.12, 19.14, and 19.15 and 19.28 yields

$$K_i = \frac{y_i}{x_i} = \frac{p_i^o}{P}\left[\frac{\phi_i^L \gamma_i \Psi_i}{\phi_i^V}\right] \tag{19.29}$$

At reasonably low pressures,

$$\frac{\phi_i^L}{\phi_i^V} \approx 1, \text{and} [\gamma_i \ \psi_i] \approx 1 \tag{19.30}$$

Equation 19.29 becomes

$$K_i = \frac{p_i^o}{P} \text{ or } y_i P = x_i p_i^o \tag{19.31}$$

19.7 Ideal Systems

The separation performance of these systems (generally under conditions of low-pressure, not close to critical point, and with similar components) can be predicted by Raoult's Law, applying to vapor and liquid in equilibrium. When one liquid is totally miscible in another, the partial pressure of each is decreased. Raoult's Law states that for any mixture, the partial pressure of any component will equal the vapor pressure of that component in the pure state multiplied by its mol fraction in the liquid mixture.

$$p_i = p_i^* \, x_i = P_i x_i \tag{19.32}$$

$$p_{ii} = P_{ii} x_{ii} \text{ (for a second component, ii, in the system)} \tag{19.33}$$

where

p_i = partial pressure, absolute, of one component in the liquid mixture
x_i = mol fraction of component, i, in the liquid mixture
$P_i^* = P_i$ = vapor pressure of component, i, in its pure state; P^*_{ii} similar by analogy

Many mixtures of liquids that do not follow Raoult's Law, since this represents the performance of ideal mixtures. For systems which do follow the ideal gas law and Raoult's Law for the liquid, for each component,

$$y_i = \frac{P_i}{\pi} = \frac{p_i^* x_i}{\pi} \tag{19.34}$$

(Raoult's Law combined with Dalton's Law)
y_i = mol fraction of component, i, in vapor
π = system total pressure

Raoult's Law is not applicable as the conditions approach critical, and for hydrocarbon mixtures accuracy is lost above about 60 psig [24]. Dalton's Law relates composition of the vapor phase to the pressure and temperature well below the critical pressure, i.e., total pressure of a system is the sum of the partial pressures of its components:

$$\pi = p_1 + p_2 + p_3 + \ldots \tag{19.2}$$

where

$p_1, p_2, \ldots$= partial pressures of components numbered 1,2, ...

Therefore, for Raoult's and Dalton's Laws to apply, the relationship between the vapor and liquid compositions for a given component of a mixture is only a function of pressure and temperature, and independent of the other components in the mixture.

For a binary system, Raoult's law states that the partial pressure in the liquid phase can be expressed as

$$p_A = x_A \, p_A^o \tag{19.35}$$

where x_A is the mole fraction of component A, and p_A^o is the vapor pressure of component A.

For the vapor phase, Dalton's law can be expressed as:

$$p_A = \pi\, y_A \tag{19.36}$$

where y_A is the mole fraction of component A. Combining Equation 19.35 and Equation 19.36 gives:

$$x_A\, p_A^o = \pi\, y_A \tag{19.37}$$

$$K = \frac{y_A}{x_A} = \frac{p_A^o}{\pi} \tag{19.38}$$

where K is the equilibrium constant.

Few multicomponent systems exist for which completely generalized equilibrium data are available. The most widely available data are those for vapor–liquid systems, frequently referred to as vapor–liquid equilibrium distribution coefficients or the K value. The K values vary with temperature and pressure, and a selectivity that is equal to the ratio of the K values is used. For vapor–liquid systems, this is referred to as the relative volatility α. For a binary system:

$$\alpha_{A-B} = \frac{K_A}{K_B} \tag{19.39}$$

where

$$K_A = y_A / x_A \tag{19.40}$$

and

$$K_B = y_B / x_B \tag{19.41}$$

$$\alpha_{A-B} = \frac{y_A \cdot x_B}{x_A \cdot y_B} \tag{19.42}$$

For binary components:

$$x_A + x_B = 1 \tag{19.43}$$

or

$$x_B = 1 - x_A \tag{19.44}$$

and

$$y_A + y_B = 1 \tag{19.45}$$

or

$$y_B = 1 - y_A \tag{19.46}$$

Substituting Equations (19.44) and (19.46) into Equation 19.42 gives

$$\alpha_{A-B} = \frac{y_A(1 - x_A)}{x_A(1 - y_A)} \tag{19.47}$$

Re-arranging Equation 19.47 in terms of y_A gives:

$$y_A = \frac{\alpha_{A-B}\, x_A}{[1 + (\alpha_{A-B} - 1)x_A]} \tag{19.48}$$

Equation 19.48 relates equilibrium compositions x_A and y_A in terms of the relative volatility. If it is assumed that α is independent of temperature and composition, Equation 19.48 becomes the equation of the equilibrium line or the x-y curve shown in Figure 19.7. This can be used to determine the compositions of vapor and liquid in equilibrium.

19.8 Henry's Law

Henry's Law applies to the vapor pressure of the solute in dilute solutions, and is a modification or Raoult's Law:

This can be stated as:

$$p_i = kx_i \tag{19.49}$$

where

p_i = partial pressure of the solute
x_i = mol fraction solute in solution
k = experimentally determined Henry's constant

Referring to Figure 19.5, Henry's Law would usually be expected to apply on the vaporization curve for about the first 1 in. of length, starting with zero, i.e., over the dilute end, whereas Raoult's Law applies to the upper end of the curve.

Carroll [25] discusses Henry's Law in detail and explains its limitations. This constant is a function of the solute-solvent pair and the temperature, but not the pressure, since it is only valid at infinite dilution.

19.8.1 Strict Henry's Law

It is equal to the reference fugacity only at infinite dilution. From [25]:

$$x_i H_{ij} = y_i P \tag{19.50}$$

for restrictions of: $x_i < 0.01$ and $P < 200$ kPa

19.8.2 Simple Henry's Law

This can be stated as:

$$x_i H_{ij} = P \tag{19.51}$$

for restrictions of: $x_i < 0.01$, $y_i = 0$, and $P < 200$ kPa

$$K = \frac{y_i}{x_i} = \frac{p_i^*}{p} \tag{19.52}$$

where

H_{ij} = Henry's constant
x_i = mol fraction of solute component, i, in liquid
P = pressure, absolute
y_i = mol fraction of solute component, i, in vapor
y_j = mol fraction solvent component, j, in vapor
kPa = metric pressure

Care must be exercised that the appropriate assumptions are made in the evaluation. This requires experience and/or experimentation. Carroll [26] presents Henry's Law constant evaluation for several multicomponent mixtures; (1) a non-volatile substance (such as a solid) dissolved in a solvent, (2) a gas in a solution of aqueous electrolytes, (3) mixed electrolytes, (4) mixed solvents, i.e., a gas in equilibrium with a solvent composed of two or more components, (5) two or more gaseous solutes in equilibrium with a single solvent, (6) complex, simultaneous phase and chemical equilibrium.

19.9 K-Factor Hydrocarbon Equilibrium Charts

Hydrocarbon systems are amongst those for which most data have been developed. See following paragraph on K-factor charts. For systems of chemical components where such factors are not developed, the basic relation is:

$$K_i = \frac{y_i}{x_i} = \frac{\gamma_i}{\phi_i} \frac{p_i^*}{P} \tag{19.53}$$

For ideal systems: $\gamma_i = \phi_i$

where

K_i = mol fraction of component, i, in vapor phase in equilibrium divided by mol fraction of component, i, in liquid phase in equilibrium
K_i = equilibrium distribution coefficient for system's component, i
P_i^* = vapor pressure of component, i, at temperature
P = total pressure of system = π
γ = activity coefficient
ϕ = fugacity coefficient

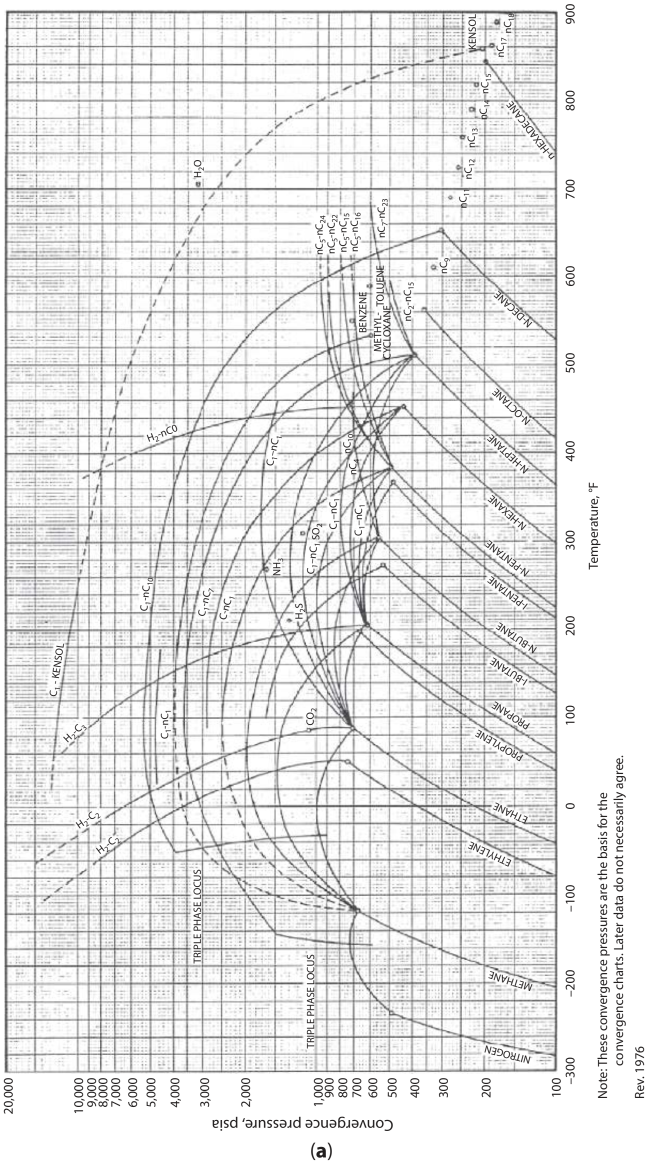
The concept of ideality is usually a good approximation for close boiling components of a system, where the components are all of the same "family" of hydrocarbons or chemicals; for example, a mixture of paraffin hydrocarbons. When "odd" or non-family components are present, or if the system has a wide boiling range the probability of deviations from non-ideality becomes greater.

Ideal conditions are often assumed for preliminary calculations, followed by more rigorous design methods. These may be completely satisfactory, particularly when the activities of the individual components approximate to 1.0.

Many process components do not however conform to the ideal gas laws, and erroneous results will be obtained if this behavior is assumed to hold. Errors are not serious when the deviation from ideal is not significant. When data are available to confirm the ideality, or non-ideality of a system, then the choice of approach is much more straightforward.

K-factors for vapor–liquid equilibrium ratios are usually associated with various hydrocarbons in the presence of some common impurities as nitrogen, carbon dioxide, and hydrogen sulfide [27]. The K-factor is (Equation 19.40) the equilibrium ratio of the mole fraction of a component in the vapor phase divided by the mole fraction of the same component in the liquid phase. K is generally considered a function of the mixture composition, in which a specific component occurs, plus the temperature and pressure of the system at equilibrium.

The Gas Processors Suppliers Association [28] provides a detailed background development of the K-factors and the use of convergence pressure (see Figures 19.11a–19.11d), i.e., the pressure of the system at a temperature when vapor-liquid separation is no longer possible. Convergence pressure alone does not represent a system's composition

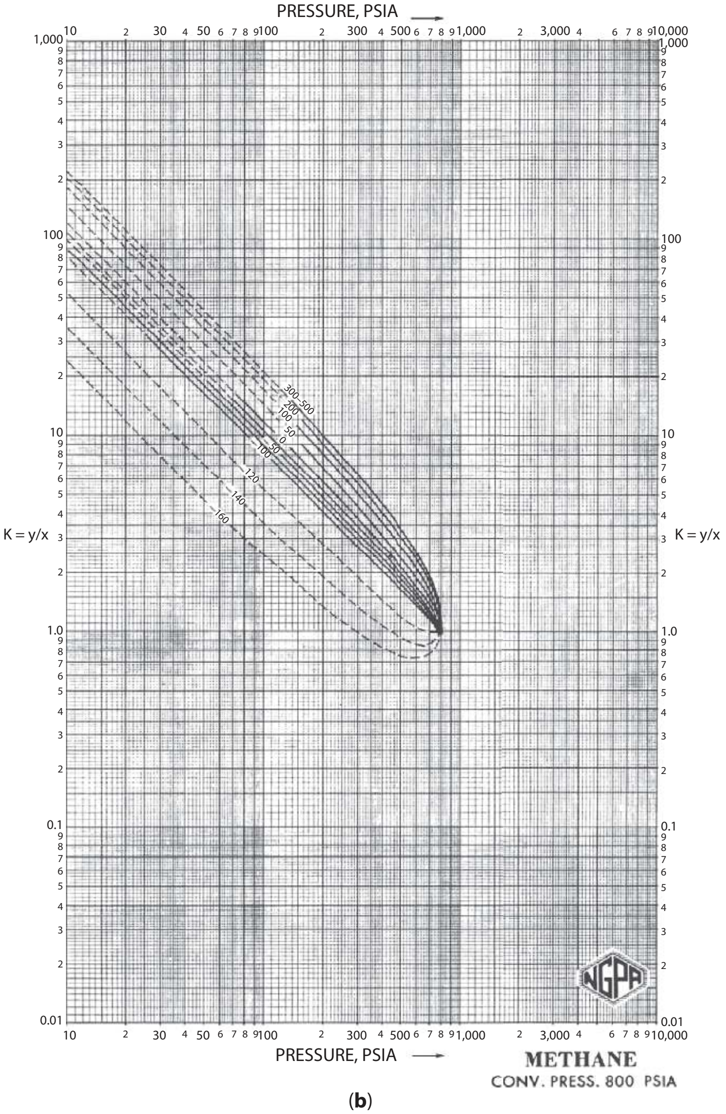


Note: These convergence pressures are the basis for the convergence charts. Later data do not necessarily agree.

Rev. 1976

(**a**)

Figure 19.11a Convergence pressures for hydrocarbons (critical locus). (Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla).



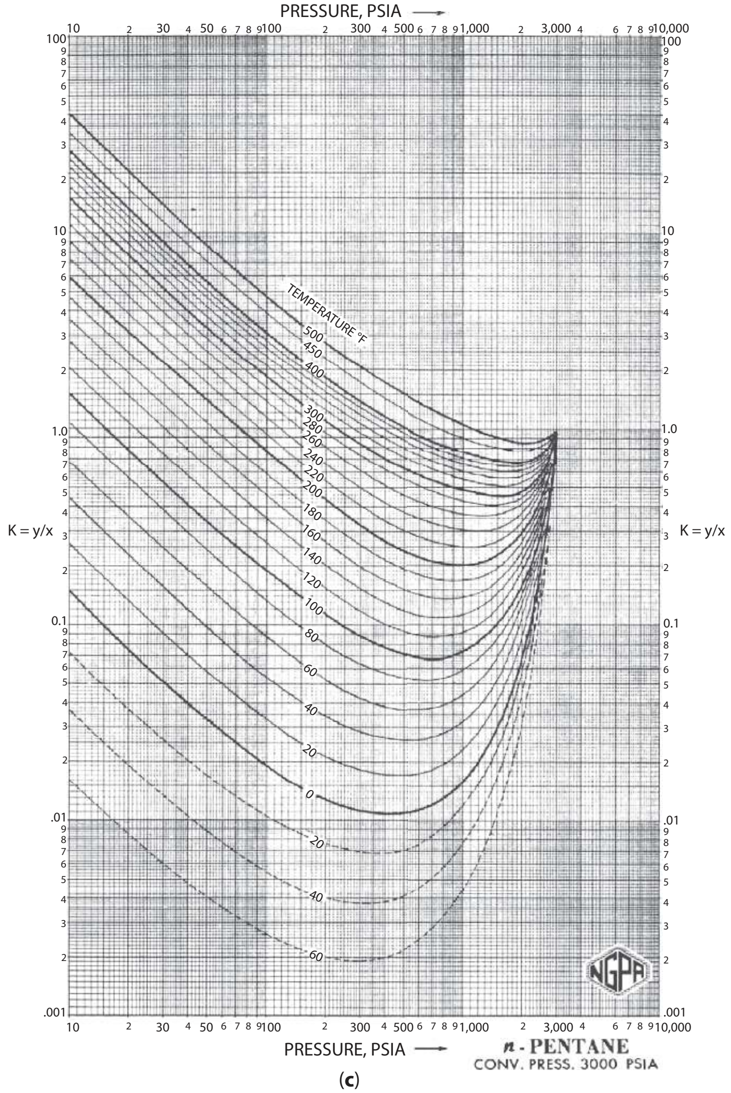
(**b**)

Figure 19.11b Pressure vs. K for methane (CH_4) at convergence pressure of 3000 psia (20,700 kPa). (Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla).

effects in hydrocarbon mixtures, but the concept does provide a rather rapid approach for systems calculations and is used for many industrial calculations. These are not well adapted for very low temperature separation systems. Figure 19.11d shows the chart of equilibrium constant for a methane–ethane binary system.

Charts of Reference [28] are for binary systems unless noted otherwise. The convergence pressure can usually predict equilibrium composition with reasonable accuracy for vapor and liquid phases around the convergence pressure and corresponding temperature. The convergence pressure is that pressure where all proportion of vapor and liquid focus to a point (converges). Generally, vapor and liquid cannot coexist at a pressure higher than the convergence pressure. The K values of the components (binary) approach 1.0. Typically, the convergence pressure is higher than the critical pressure of either of the pure components [29].

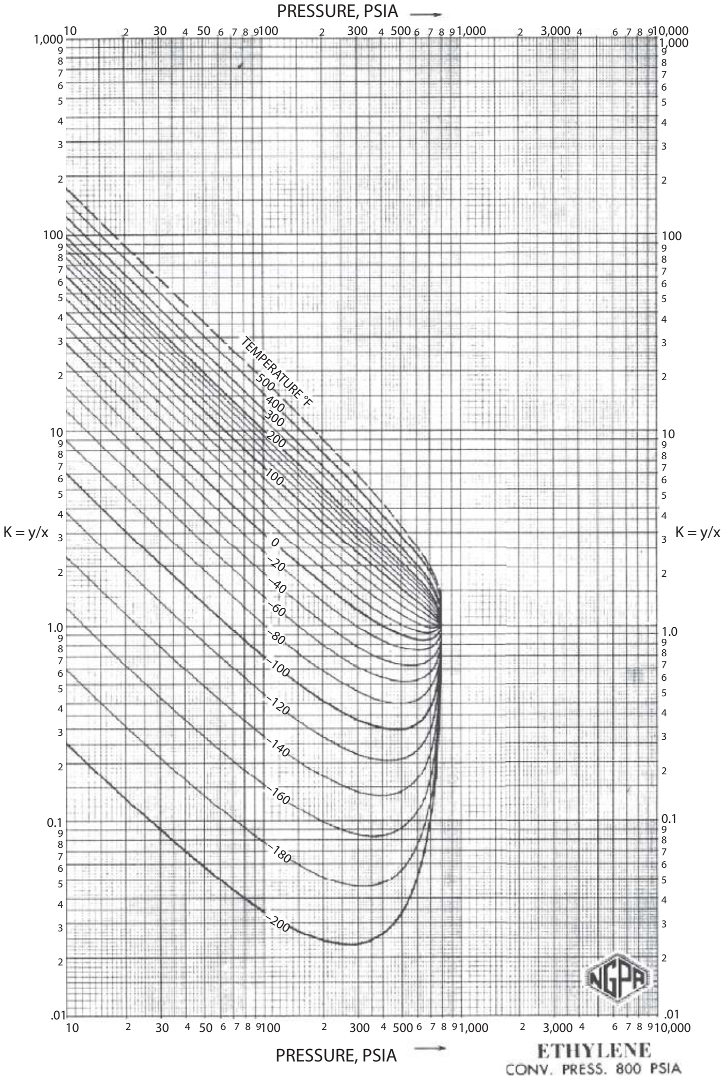
The convergence pressure generally is a function of the liquid phase, and assumes that the liquid composition is known from a flash calculation using a first estimate for convergence pressure, and is usually the critical pressure of a system at a given temperature. The following procedure is recommended by Reference [28]:



(c)

Figure 19.11c Pressure vs. K for n-pentane (n-C_5H_{12}) at convergence pressure of 3000 psia (20,700 kPa). (Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla).

Step 1. Assume the liquid phase composition or make an approximation. (If there is no guide, use the total feed composition).
Step 2. Identify the lightest hydrocarbon component that is present at least 0.1 mole % in the liquid phase.
Step 3. Calculate the weight average critical temperature and critical pressure for the remaining heavier components to form a pseudo binary system. (A shortcut approach good for most hydrocarbon systems is to calculate the weight average T_c only).
Step 4. Trace the critical locus of the binary mixture consisting of the light component and pseudo heavy component. When the averaged pseudo heavy component is between two real hydrocarbons, an interpolation of the two critical loci must be made.
Step 5. Read the convergence pressure (ordinate) at the temperature (abscissa) corresponding to that of the desired flash conditions, from Figure 19.11a [28].
Step 6. Using the convergence pressure determined in Step 5, together with the system temperature and system pressure, obtain K-values for the components from the appropriate convergence-pressure K-charts.
Step 7. Make a flash calculation with the feed composition and the K-values from Step 6.
Step 8. Repeat Steps 2 through 7 until the assumed and calculated convergence pressures agree within an acceptable tolerance.



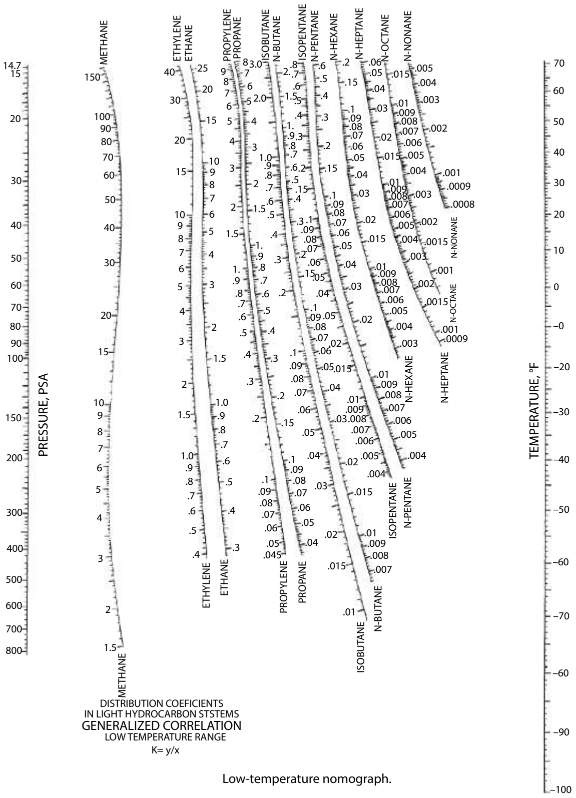
(**d**)

Figure 19.11d Pressure vs. K for ethylene (C_2H_4) at convergence pressure of 800 psia. (Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla).

For n-pentane at convergence pressure of 3000 psia (nearest chart), the K-value reads 0.99 (see Figure 19.11c) and other K-values are shown in Appendix F. The DePriester charts [30] provide a reasonable check on this (Figures 19.12a–d). Interpolation between charts of convergence pressure can be calculated, depending on how close the operating pressure is to the convergence pressure. The K-factor charts (or K-values) do not change rapidly with convergence pressure. P_k (psia) [28].

For light hydrocarbons, the approximate K values can be determined from the nomographs prepared by DePriester as illustrated in Figures 19.12a–d for different temperature ranges. If temperature and/or pressure of the equilibrium mixture are unknown, a trial-and-error method is carried out. Perry and Green [16], and Smith and Van Ness [18] provide DePriester charts in other temperature and pressure units. McWilliams [31] has developed an equation to fit DePriester charts and is represented by

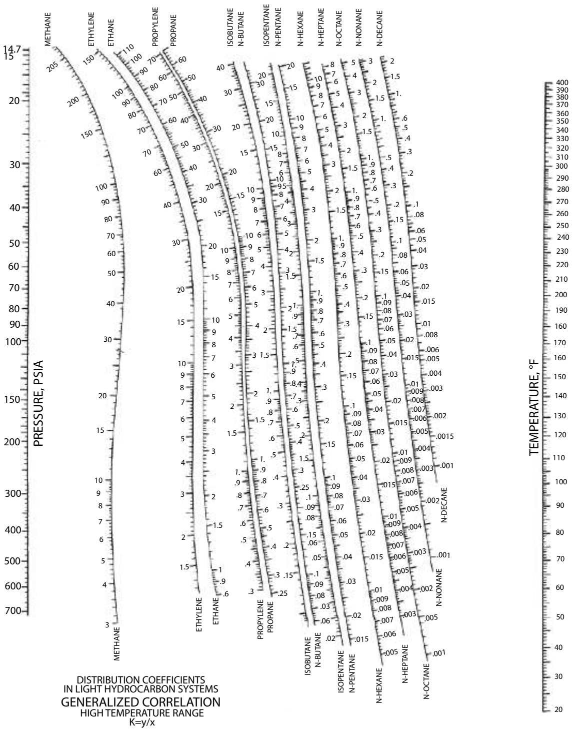
$$\ln K = a_{T1}/T^2 + a_{T2}/T + a_{T6} + a_{pl}\ln p + a_{p2}/p^2 + a_{p3}/p \tag{19.54}$$



Low-temperature nomograph.

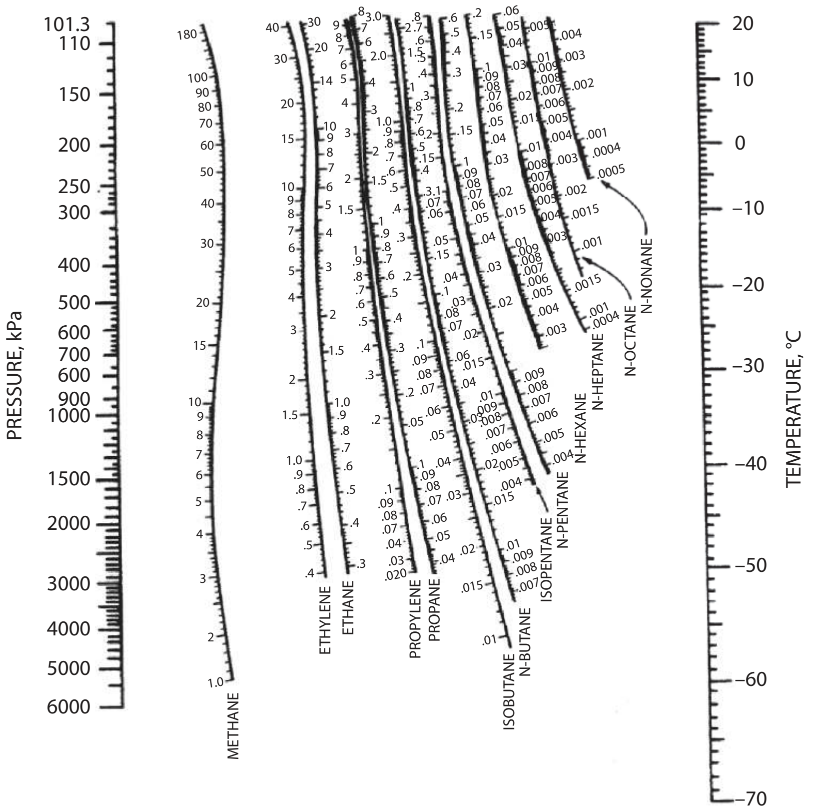
(a)

Figure 19.12a DePriester Charts; K-values of light hydrocarbon systems, generalized correlations, low-temperature range. (Used by permission, DePriester, C. L., The American Institute of Chemical Engineers, Chemical Eng. Prog. Ser., 49, No. 7 (1953). All rights reserved).



(b)

Figure 19.12b DePriester Charts; K-values of light hydrocarbon systems, generalized correlations, high-temperature range. (Used by permission, DePriester, C. L., The American Institute of Chemical Engineers, Chemical Eng. Prog. Ser., 49, No. 7 (1953). All rights reserved).



(c)

Figure 19.12c Modified DePriester Chart (in S.I. units) at low temperature (D.B., Dadyburjor, Chem. Eng. Prog., 85, April 1978; copyright 1978, AIChE; reproduced by permission of the American Institue of Chemical Engineers).

where

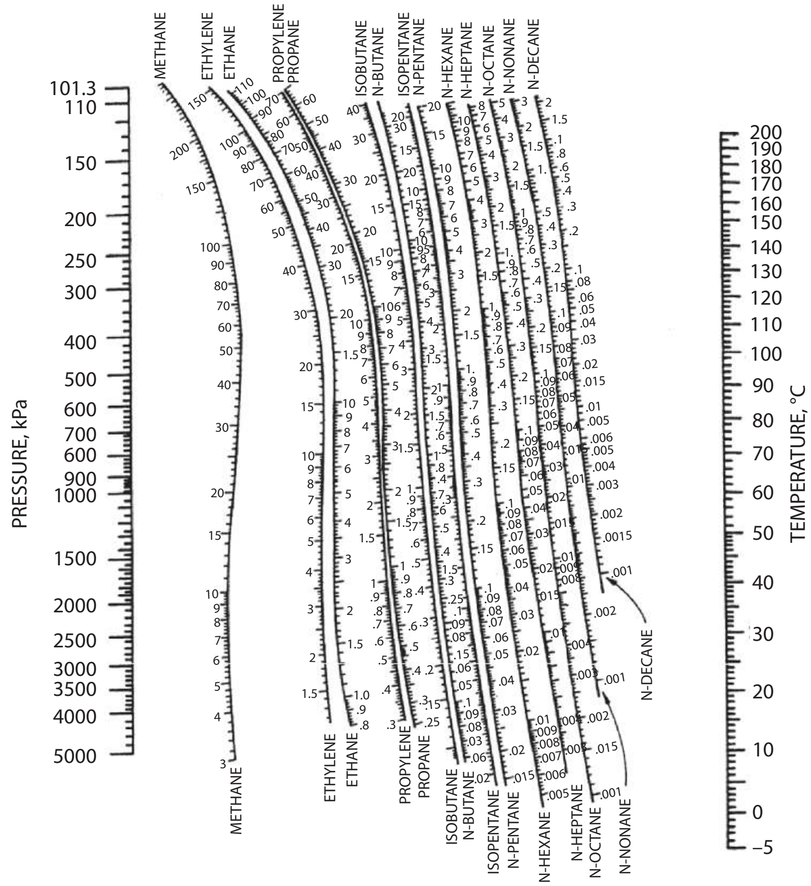
a_{T1}, a_{T2}, a_{T6}, a_{p1}, a_{p2}, a_{p3} = constants
K = equilibrium constant
p = absolute pressure, psia
T = absolute temperature (°R = °F + 460).

Equation 19.54 is valid from 365.7°R (−70°C) to 851.7 °R (200°C) and from pressures from 14.69 psia (101.3 kPa) to 870.1 psia (6000 kPa). If K and p are known, Equation 19.54 can be solved for T. The relationship values can be easily incorporated into a computer program and can be determined more accurately than from the charts. Table 19.2 shows the constants for various components and Equation 19.54 can be simplified for all components except n-octane and n-decane. This is expressed by:

$$\ln K = a_{T1}/T^2 + a_{T6} + a_{p1}\ln p \tag{19.55}$$

Solving for T from Equation 19.54 gives

$$T = \left[\frac{a_{T1}}{\left(\ln K - a_{T6} - a_{p1}\ln p - \dfrac{a_{p2}}{p^2} - \dfrac{a_{p3}}{p}\right)}\right]^{0.5} \tag{19.56}$$



(**d**)

Figure 19.12d Modified DePriester Chart (in S.I. units) at high temperature (D.B., Dadyburjor, Chem. Eng. Prog., 85, April 1978; copyright 1978, AIChE; reproduced by permission of the American Institue of Chemical Engineers).

The K values are used along with the stoichiometric equations which state that the sum of the mole fractions in liquid and vapor phases must equal to unity as defined by Equations 19.43 and 19.45, respectively. For N number of components, Equations 19.43 and 19.45 can be expressed by

$$\sum_{i=1}^{N} x_i = 1.0, \quad \sum_{i}^{N} y_i = 1.0 \tag{19.57}$$

where

N = number of components.

If only one component is present, then x = 1.0 and y = 1.0, which implies that $K_i = y/x = 1.0$. This is a simple way of determining the boiling point of a pure compound at any pressure. For example, to determine the boiling point of propane at p = 200 psia, place a straightedge on p = 200 and K = 1.0 on the propane scale on Figure 19.12a. The temperature on the right side of the scale is T = 100 °F (560 °R) as the boiling point.

Using Equation 19.54 and values from Table 19.2 gives

$$\ln K = a_{T1}/T^2 + a_{T2}/T + a_{T6} + a_{p1}\ln p + a_{p2}/p^2 + a_{p3}/p \tag{19.54}$$

Table 19.2 Constants for fit to K values using Equation 19.54.

Compound	a_{T1}	a_{T2}	a_{T6}	a_{p1}	a_{p2}	a_{p3}	Mean error
Methane	-292,860	0	8.2445	-0.8951	59.8465	0	1.666
Ethylene	-600,076.875	0	7.90595	-0.84677	42.94594	0	2.65
Ethane	-687,248.25	0	7.90694	-0.88600	49.02654	0	1.95
Propylene	-923,484.6875	0	7.71725	-0.87871	47.67624		1.90
Propane	-970,688.5625	0	7.15059	-0.76984	0	6.90224	2.35
Isobutane	-1,166,846	0	7.72668	-0.92213	0	0	2.52
n-Butane	-1,280,557	0	7.94986	-0.96455	0	0	3.61
Isopentane	-1,481,583	0	7.58071	-0.93159	0	0	4.56
n-Pentane	-1,524,891	0	7.33129	-0.89143	0	0	4.30
n-Hexane	-1,778,901	0	6.96783	-0.84634	0	0	4.90
n-Heptane	-2,013,803	0	6.52914	-0.79543	0	0	6.34
n-Octane	0	-7646.81641	12.48547	-0.73152	0	0	7.58
n-Nonane	-2,551,040	0	5.69313	-0.67818	0	0	9.40
n-Decane	0	-9760.45703	13.80354	-0.71470	0	0	5.69

Note: T is in °R, and p is in psia.
(Source: McWilliams, M. L., "An Equation to Relate K-factors to Pressure and Temperature", Chem. Eng., 80 (25), 138, Oct 29, 1973).

and

$$a_{T1} = -970{,}688.5625,\ a_{T2} = 0,\ a_{T6} = 7.15059,\ a_{p1} = -0.76984,\ a_{p2} = 0,\ a_{p3} = 6.90224$$

$$\ln(1) = \frac{-970{,}688.5625}{T^2} + 0 + 7.15059 - 0.76984\ln(200) + \frac{0}{200^2} + \frac{6.90224}{200}$$

$$T = 558.98\ °R\ (98.98°F)$$

This gives a 1.03% deviation from DePriester chart.

The K-factor charts represent pure components and pseudo binary systems of a light hydrocarbon plus a calculated pseudo heavy component in a mixture, when several components are present. For these more complex mixtures, it is necessary to determine the average molecular weight of the system on a methane-free basis, and then interpolate the K-value between the two binaries whose heavy component lies on either side of the pseudo-components. If nitrogen is present by more than 3–5 mol%, the accuracy becomes poor. Reference [28] provides a more detailed explanation and a more complete set of charts.

19.10 Non-Ideal Systems

A system is non-ideal because of the interaction between molecules alters its behavior. The following groups all contribute to non-ideality; hydroxyl groups (–OH), ketone groups (–C=O), aldehyde groups (CHO), halogens

(–Cl, –Br), and carboxylic acid groups (–COOH). Systems of two or more hydrocarbons, chemical, and water components may be non-ideal for a variety of reasons. Accurate experimental data are necessary to predict the distillation performance of these systems. The use of empirical relationships to predict vapor pressure/concentration relationships at specific temperatures and pressures is inaccurate at best.

Prausnitz [32] presents a thorough analysis of the application of empirical techniques in the absence of experimental data. The basic question is to determine accurately the distribution of the respective system components between the liquid and gaseous phases. The concepts of fugacity and activity are fundamental to this. For a pure ideal gas, the fugacity is equal to the pressure, and for a component, i, in a mixture of ideal gases it is equal to its partial pressure y_i P, where P is the system pressure. As the system pressure approaches zero, the fugacity approached ideal. For many systems the deviations from unity are minor at system pressures less than 25 psig.

The ratio f/f° is termed activity, a. [**Note**: This is not the activity coefficient.] The activity is an indication of how "active" a substance is relative to its standard state (not necessarily zero pressure), f°. The standard state is the reference condition; most references are to constant temperature, with composition and pressure varying as required. Fugacity becomes a corrected pressure, representing a specific component's deviation from ideal. The fugacity coefficient is:

$$\phi_i = \frac{f_i}{y_i P} \tag{19.58}$$

The Virial Equation of State for gases is generally:

$$Z = \frac{PV}{RT} = 1 + \frac{B}{V} + \frac{C}{V^2} + \frac{D}{V^3} + \ldots \tag{19.59}$$

where

B, C, D, etc. = Virial coefficients, independent of pressure or density, and for pure components are functions of temperature only.
V = molar volume.
Z = compressibility factor.

Fugacities and activities can be determined using this concept. Other important equations of state which can be related to fugacity and activity have been developed by Redlich–Kwong [33] with Chueh [34], which develops the original Redlich–Kwong equation, and Palmer's summary of activity coefficient methods [35].

Activity coefficients are equal to 1.0 for an ideal solution when the mole fraction is equal to the activity. The activity (a) of a component, i, at a specific temperature, pressure and composition is defined as the ratio of the fugacity of i at these conditions to the fugacity of i at the standard state [32].

$$a(T,P,x) = \frac{f_i(T,P,x)}{f_i(T^o,P^o,x^o)}, \text{liquid phase} \tag{19.60}$$

(Zero superscript indicates a specific pressure and composition)

The activity coefficient γ_i is

$$\gamma_i = \frac{a_i}{x_i} = 1.0 \text{ for ideal solution} \tag{19.61}$$

The ideal solution law, Henry's Law, also enters into the establishment of performance of ideal and non-ideal solutions.

The Redlich–Kister [36, 37] equations provide a good technique for representing liquid phase activity and classifying solutions.

The Gibbs–Duhem equation allows the determination of activity coefficients for one component from data for those of other components.

Wilson's [38] equation has been found to be quite accurate in predicting the vapor-liquid relationships and activity coefficients for miscible liquid systems. The results can be expanded to as many components in a multicomponent system as necessary without additional data requiring. This makes Renon's and Prausnitz's [39] techniques valuable for multicomponent systems when applying computational solutions.

Renon's technique is applicable to partially miscible systems as well as those with complete miscibility. This is described in the reference above and in Reference [32].

Other specific techniques applicable to particular situations; they should be investigated to make sure that they apply well to the system under consideration. They are often expressed in terms as the effective "K" for the components, i, of a system. Frequently used methods are those of Chao–Seader (CS), Peng–Robinson (PR), Renon, Redlich–Kwong, Soave Redlich–Kwong (SRK/RSK), Wilson. Table 19.3 lists the equations of state that are applied most widely.

Table 19.3 Useful equations of state.

Name	Equation	Equation constants and functions
(1) ideal gas law	$P \quad \frac{RT}{V}$	None
(2) Generalized	$P \quad \frac{ZRT}{V}$	Z =Z (P_r, T_r, Z_c or ω) as derived from data
(3) van-der-Waals	$P = \frac{RT}{(V-b)} - \frac{a}{V^2}$	a and b are species dependent constants and estimated from the critical temperature and pressure.
(4) Redlich-Kwong (R-K)	$P = \frac{RT}{V-b} - \frac{a}{V^2 + bV}$	$b = 0.08664RT_c/P_c$ $a = 0.42748R^2T_c^{2.5}/P_cT^{0.5}$
(5) Soave-Redlich-Kwong (S-R-K or R-K-S)	$P = \frac{RT}{V-b} - \frac{a}{V^2 + bV}$	$b = 0.08664RT_c/P_c$ $a = 0.42748R^2T_c^2[1 + f_\omega(1-T_r^{0.5})]^2/P_c$ $f_\omega = 0.48 + 1.574\omega - 0.176\omega^2$
(6) Peng-Robinson (P-R)	$P = \frac{RT}{V-b} - \frac{a}{V^2 + 2bV - b^2}$	$b = 0.07780RT_c/P_c$ $a = 0.45724R^2T_c^2[1 + f_\omega(1 - T_r^{0.5})]^2/P_c$ $f_\omega = 0.37464 + 1.54226\omega - 0.26992\omega^2$

The van-der-Waals equation (3) in Table 19.3 was the first successful approach to the formulation of an equation of state for non-ideal gas. It is seldom used because its range of application is too narrow. The constants a and b are estimated from the critical temperature T_c and pressure P_c. The generalized equation of state (2) given in Table 19.3 defines the compressibility factor Z, which is a function of reduced pressure P_r, $(P_r = P/P_c)$ and reduced temperature T_r, $(T_r = T/T_c)$. The critical compressibility factor Z_c, or the acentric factor, ω developed by Pitzer *et al.* [40] is determined from experimental P-V-T data. It accounts for differences in molecular shape, is determined from the vapor pressure curve and defined as

$$\omega = \left[-\log\left(\frac{P^s}{P_c}\right)_{T_r = 0.7} \right] - 1.0 \tag{19.62}$$

Poling *et al.* [41] give extensive lists of values of ω. The equation developed by Redlich and Kwong (4) in Table 19.3 is a considerable improvement over the van der Waals equation. Shah and Thodos [42] showed that the simple R-K equation gives comparable accuracy with equations containing many more empirical constants. The R-K equation can also approximate the liquid phase region. A cubic equation in V is obtained if the R-K equation is expanded to obtain a common denominator. Alternatively, (3) and (4) in Table 19.3 can be combined to eliminate V, and hence give the compressibility factor Z form of the R-K equation, and is expressed by:

$$Z^3 - Z^2 + (A - B - B^2)Z - AB = 0 \tag{19.63}$$

where

$$A = \frac{aP}{R^2 T^2} \tag{19.64}$$

$$B = \frac{bP}{RT} \tag{19.65}$$

Equation 19.63 can be solved analytically for three roots.

19.11 Thermodynamic Simulation Software Programs

One of the basic tests of a thermodynamic model is to find relevant experimental data to test the predictions of the model. The comparison helps to decide whether the model is good or whether another type would be more suitable. Simulation programs have in-built packages which provide accurate thermodynamic property predictions for hydrocarbons and non-hydrocarbons fluids. Many of these simulation programs have databases of over 1500 components and over 10,000 of fitted binary coefficients. If a component cannot be found within the database, the software programs provide a selection of estimation methods for creating appropriate hypothetical compounds (e.g., pseudo hydrocarbon components of crude oil for determining the true boiling point (TBP) of crude oil). Experimental pure component data can be used as input to the regression package. Alternatively, the designers can supply a set of their own data. The regression package will fit the input data to one of the mathematical expressions, and thus allows the designers to obtain simulation results for specific thermophysical properties which closely match their experimental data.

Selecting the correct thermodynamic models from packages in the simulation programs (e.g., UniSim, Hysys, Aspen, ChemCad, Pro/II, ProSim, etc.) enables the designer to predict properties of mixtures ranging from well-defined light hydrocarbon systems to complex oil mixtures and highly non-ideal (non-electrolytic) chemical systems. To obtain a successful simulation, the input parameters and regression model must be carefully chosen. Changes in temperature and pressure can affect the accuracy of a simulation quite drastically if inappropriate choices are made.

The thermodynamic models used to predict the chemical interaction properties are the NRTL, Wilson, SRK, UNIQUAC, and UNIFAC models. These are incorporated in various simulation packages such as Pro II, HYSYS, Aspen Plus, ProSim Plus, ChemCad and UniSim (Honeywell).

The procedure for selection, as illustrated by Seader and Henley [43] is as follows:

RK	= Redlich–Kwong
PR	= Peng Robinson
SRK/RSK	= Soave–Redlich–Kwong
NRTL	= Non-random two-liquid
UNIQUA	= Universal quasichemical
UNIFAC	= UNIQUAC Functional group activity

1. Characterize the mixture by chemical types present: Light gases (LG), Hydrocarbons (HC), Polar organic compounds (PC), and aqueous solutions (A), with or without Electrolytes (E).
2. If the mixture is (A) with no (PC), then if electrolytes are present, select the modified NRTL equation. Otherwise, select a special model, such as one for sour water (containing NH_3, H_2S, CO_2, etc.) or aqueous amine solutions.
3. If the mixture contains (HC), with or without (LG), and has a wide boiling range, choose the corresponding states method of Lee–Kesler–Plöcker. If the boiling range of a mixture of (HC) is not larger, selection depends on the pressure and temperature.
4. For all temperatures and pressures, the Peng–Robinson (PR) equation is suitable.
5. For all pressures and non-cryogenic temperatures, the Soave–Redlich–Kwong (SRK/RSK) equation is recommended.
6. For all temperatures, but not pressures in the critical region, the Benedict–Webb–Rubin–Starling method is suitable.
7. If the mixture contains (PC), the selection depends on whether (LG) are present. If they are, the Predictive Soave–Redlich–Kwong (PSRK) model is recommended. If not, then a suitable liquid-phase activity coefficient method is chosen.
8. If the binary interaction coefficients are not available, select the UNIFAC method. Its results should be considered as only a first approximation.
9. If the binary interaction coefficients are available and splitting in two liquid phases will not occur, select the Wilson or NRTL equation.
10. If phase splitting is likely, select the NRTL or UNIQUAC model.

Table 19.4 shows the conditions of some of these models, and Figure 19.13 [19] gives the selection process in the form of a flow chart. Table 19.5 lists some typical systems and recommended correlations.

Table 19.4 Approximate guides for selection of K-values methods.

Chemical systems		
Low MW Alcohol and Hydrocarbons		Wilson
Higher MW Alcohol and Hydrocarbons		NRTL
Hydrogen Bonding Systems		Margules
Liquid-Liquid Equilibrium		NRTL/UNIQUAC
Water as a Second Liquid Phase		NRTL
Components in a Homologous Family		UNIQUAC
Low Pressure Systems with Associating Vapor Phase		
Light Hydrocarbon and Oil Systems		
Natural Gas Systems w/sweet and sour gas		SRK/PR
Cryogenic Systems		SRK/PR
Refinery Mixtures with $p < 5000$ psia		SRK/PR
Hydrotreaters and Reformers		Grayson-Stread
Simple Paraffinic Systems		SRK/PR
Heavy Components w/NBP > 1,000°F		BK10
Based on Polarity and Ideality		
nonpolar - nonpolar	ideal & non-ideal	any activity coefficient model
nonpolar – weakly polar	ideal	any activity coefficient model
nonpolar – weakly polar	non-ideal	UNIQUAC
nonpolar – strongly polar	ideal	UNIQUAC
nonpolar – strongly polar	non-ideal	Wilson
weakly polar – weakly polar	ideal	NRTL
weakly polar – weakly polar	non-ideal	UNIQUAC
weakly polar – strongly polar	ideal	NRTL
weakly polar – strongly polar	non-ideal	UNIQUAC
strongly polar – strongly polar	ideal	UNIQUAC

(*Continued*)

Table 19.4 Approximate guides for selection of K-values methods. (*Continued*)

Based on Polarity and Ideality		
strongly polar – strongly polar	non-ideal	NO RECOMMENDATION
aqueous – strongly polar		UNIQUAC

Key: NRTL = non-random two liquid model; SRK = Soave-Redlich-Kwong; PR = Peng-Robinson; BK10 = Braun K10 for petroleum; UNIQUAC = universal quasi-chemical theory.
(Source: Wankat, P. C., Separation Process Engineering, 2nd, Prentice Hall, 2007).

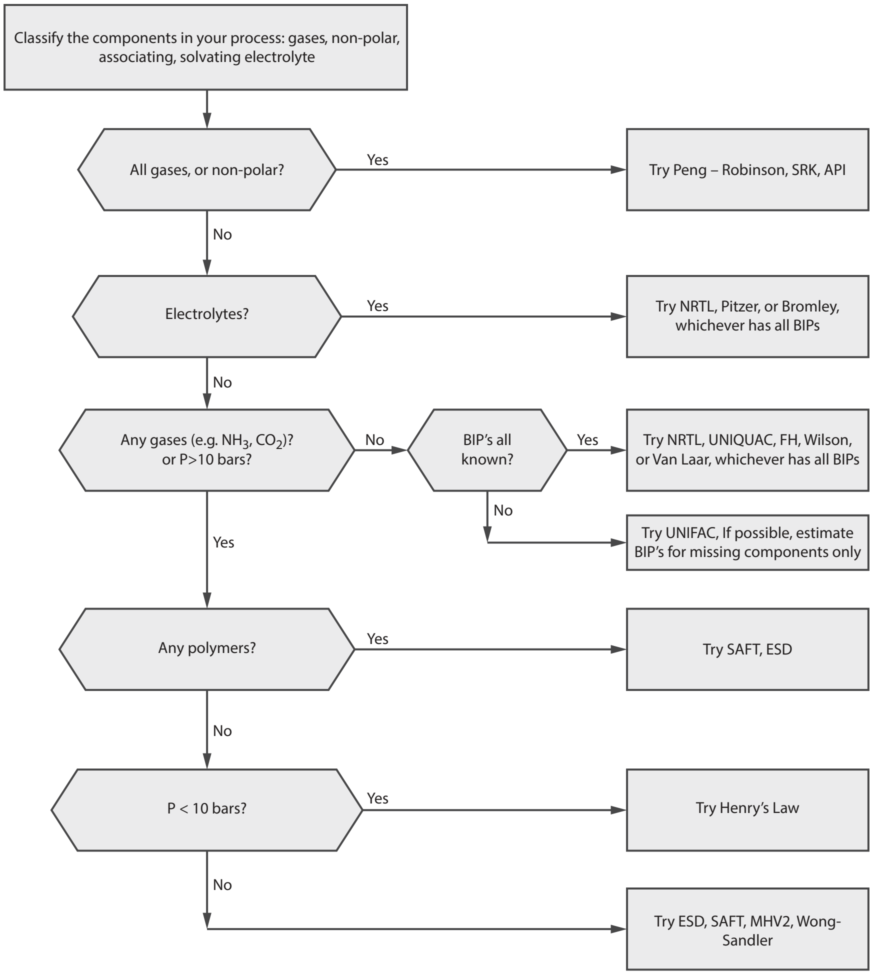


Figure 19.13 Flow chart to select the best thermodynamic model. The abbreviation BIP is used to mean binary interaction parameters (source: Elliot, J.R. and Carl T. Lira, Introductory Chemical Engineering Thermodynamics, Prentice Hall Int. Series, 1999).

19.12 Vapor Pressure

If the vapor and liquid of a pure component are in equilibrium, then the equilibrium pressure is known as the vapor pressure. Vapor pressure can be defined as the pressure that a pure component exerts at a given temperature when both liquid and vapor phases are present. It is a physical property of a pure chemical component, and depends only on temperature.

Table 19.5 Generalized Antoine constant functions for the SRK equation.

$0.6 \le T_r \le 0.7$
$A' = 4.0434 + 2.1456f - 0.43477f^2 + 0.38481f^3 - 0.050479f^4$
$B' = 4.3755 + 1.2981f - 0.53208f^2 + 0.44506f^3 - 0.074005f^4$
$C' = 0.0708033 - 0.240584f + 0.080150f^2 - 0.00990303f^3 - 0.000160472f^4$
$0.7 \le T_r \le 0.84$
$A' = 4.4401 + 2.2128f - 0.5351f^2 + 0.17368f^3 - 0.018512f^4$
$B' = 5.0075 + 1.249f - 0.78155f^2 + 0.32010f^3 - 0.047601f^4$
$C' = 0.124652 - 0.273702f + 0.0750076f^2 - 0.0137818f^3 + 0.00145038f^4$
$0.84 \le T_r \le 1.0$
$A' = 4.8817 + 2.5164f - 0.53557f^2 + 0.09639f^3 - 0.012108f^4$
$B' = 5.8938 + 1.6311f - 0.91992f^2 + 0.24694f^3 - 0.034344f^4$
$C' = 0.206497 - 0.281653f + 0.0716095f^2 - 0.0174537f^3 + 0.00236311f^4$

A', B', C' = constants in generalized Antoine equation, f = acentric factor function,
T_r = reduce temperature
(Source: DeDoes, A. J, *et al.,* CEP, pp 39-44, Jan. 2007)

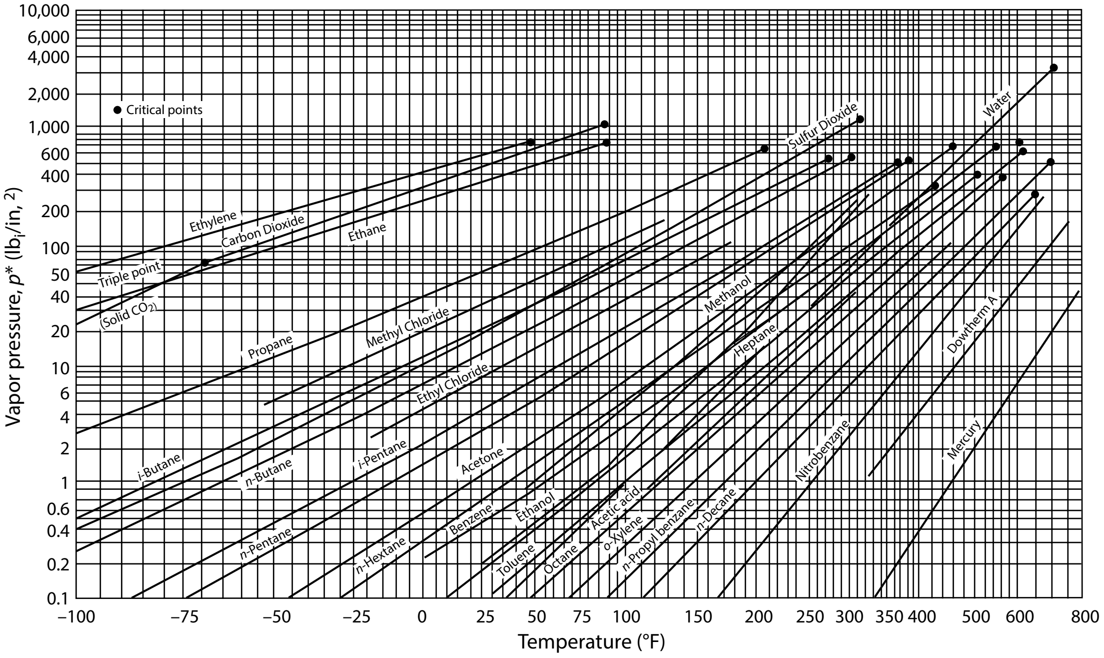
Vapor pressure is an essential parameter in the analysis of separation processes such as flash separation, distillation or absorption. Various approaches such as tables and charts, listing vapor pressure or boiling point among their physical properties are used in these analyses. These include steam tables, thermodynamic property tables and phase equilibrium diagrams such as pressure–enthalpy, pressure–volume or detailed some temperature–entropy diagrams.

DeDoes *et al.* [44] have detailed some drawbacks of these approaches. They have developed a generalized method, which combines pressure–volume–temperature (PVT) equations of state with thermodynamic criteria that are related to chemical potential or fugacity. The method uses three constants such as critical temperature, critical pressure and acentric factor. Other phase properties such as molar volume or density, residual enthalpy and residual entropy are also predicted, and generate mixture compositions, K-values or relative volatilities when combined with

Table 19.6 Generalized Antoine constant functions for the PR equation.

$0.7 \le T_r \le 0.84$
$A' = 4.8512 + 2.3617f - 0.45894f^2 - 0.070461f^3 + 0.10809f^4$
$B' = 5.5675 + 1.3213f - 0.80186f^2 + 0.11081f^3 + 0.087503f^4$
$C' = 0.145395 - 0.287367f + 0.0815004f^2 - 0.0253619f^3 + 0.00813261f^4$
$0.84 \le T_r \le 1.0$
$A' = 5.3988 + 3.0095f - 0.68621f^2 + 0.10190f^3 - 0.0027322f^4$
$B' = 6.7141 + 2.3263f - 1.3630f^2 + 0.35003f^3 - 0.035704f^4$
$C' = 0.243432 - 0.260581f + 0.0448442f^2 -0.00718139f^3 + 0.00122297f^4$

A', B', C' = constants in generalized Antoine equation, f = acentric factor function, T_r = reduce temperature.
(Source: DeDoes, A. J, *et al.*, CEP, pp 39-44, Jan. 2007)



Conversions: °F = 1.8°C + 32, 1lb/in^2 = 0.06895 bar

Figure 19.14 Cox chart vapor pressure plots. (Source: A. S. Foust *et al.*, Principles of Unit Operations, Wiley, New York, p. 550, 1960).

appropriate mixing rules or a liquid solution model. The generalized method uses a three-constant Antoine vapor pressure relationship for any substance for which the critical temperature, critical pressure and acentric factor are known. Vapor pressure predictions have been found to agree to within **±0.1%** with those generated by an algorithm-based approach over a range of reduced temperatures between 0.60 to 1.00 for Soave–Redlich–Kwong (SRK) and between 0.7 and 1.00 for Peng–Robinson (PR) as shown in Table 19.6. Figure 19.14 shows Cox chart plots of vapor pressure for various substances, Figures 19.14a and Figure 19.14b show the vapor pressures for light hydrocarbons at a low temperature range and at a high temperature range, respectively.

Vapor Pressure Determination Using the Clausius–Clapeyron and the Antoine Equations

Standard chemical engineering texts [17, 21, 45] provide equations to predict vapor-pressure of compounds, and a commonly given equation for the vapor is:

$$\log_{10} P_v = A + \frac{B}{T} + C\log_{10} T + DT + ET^2 \tag{19.66}$$

where P is the vapor pressure in mm Hg, and A, B, C, D, and E are constants. Sometimes, it is useful to estimate the saturated vapor pressure from the *Clausius–Clapeyron* equation as

$$\frac{dP^{vap}}{dT} = \frac{P^{vap}\Delta H_{vap}}{RT^2} \text{ or } \frac{d\ln P^{vap}}{dT} = \frac{\Delta H_{vap}}{RT^2} \tag{19.67}$$

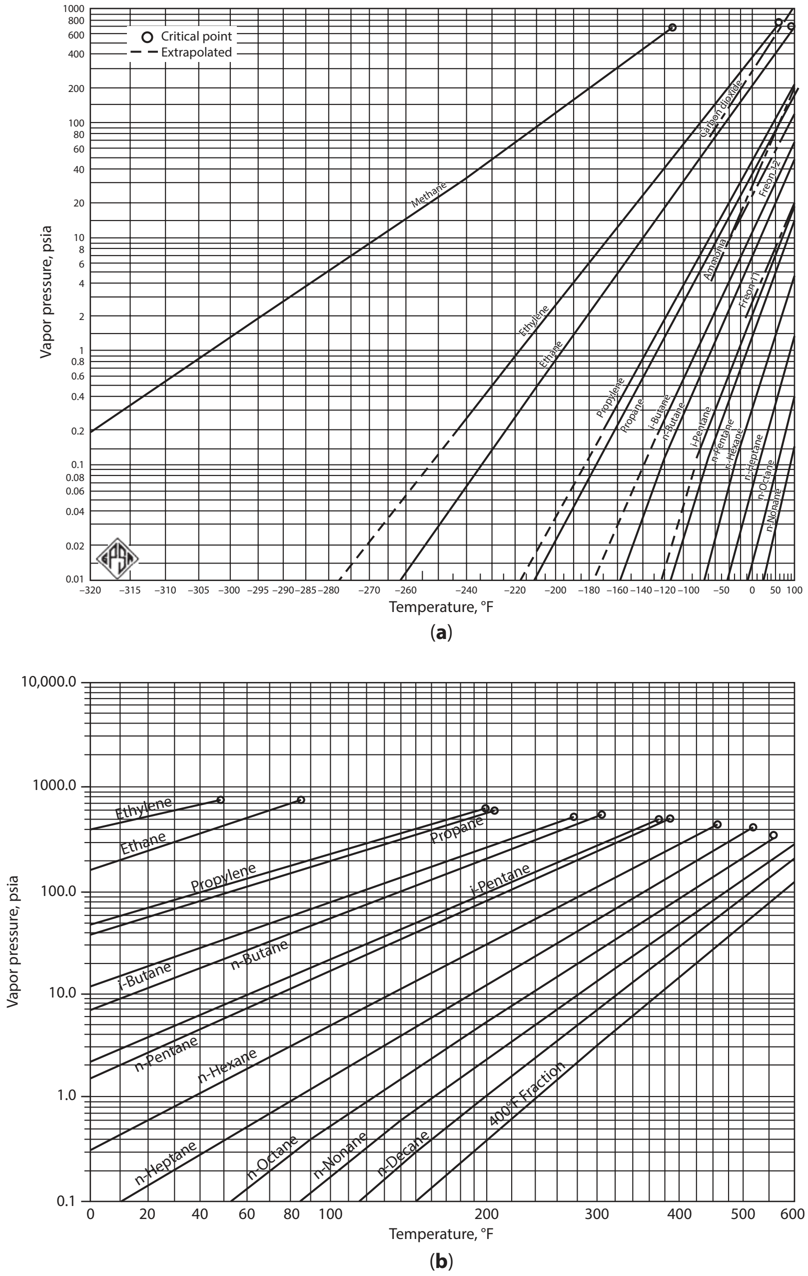


Figure 19.14 (a) Low-temperature vapor pressure for light hydrocarbons. (b) High-temperature vapor pressure for light hydrocarbons. (Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004) Tulsa, Okla).

and

$$\ln\left(\frac{P^{vap}(T_2)}{P^{vap}(T_1)}\right)=\int_{T_1}^{T_2}\frac{\Delta H_{vap}}{RT^2}dT \tag{19.68}$$

where ΔH_{vap} is the molar latent heat of vaporization and T is the absolute temperature. $\Delta H_{vap} = H^V - H^L$ is a function of temperature; but if it is assumed to be independent of temperature, Equation 19.67 can be integrated to give:

$$\ln\left(\frac{P^{vap}(T_2)}{P^{vap}(T_1)}\right)=-\frac{\Delta H_{vap}}{R}\left(\frac{1}{T_2}-\frac{1}{T_1}\right) \tag{19.69}$$

Equation 19.69 has been found to be fairly accurate for correlating the temperature dependence of the vapor pressure of liquids over limited temperature ranges.

Note: ln P^{vap} should be linear function of 1/T, where T is the absolute temperature with the lower limit of integration as the normal boiling conditions, i.e., $T_1 = T_{nb}$ at $P^{vap}(T_1) = P_{atm}$ where P_{atm} is the atmospheric pressure, ΔH_{vap} and R are in a consistent set of units. One can obtain an estimate of the temperature variation of the heat of vaporization by noting that the integration of Equation 19.69 can be carried out as an indefinite rather than definite integral, which gives:

$$\ln P^{vap}(T)=-\frac{\Delta H_{vap}}{RT}+C \tag{19.70}$$

where C is a constant. By plotting ln P^{vap} versus 1/T, we should get a straight line with a slope equal to $-\Delta H_{vap}/R$, if the heat of vaporization is independent of temperature, and a curve if $-\Delta H_{vap}/R$ varies with temperature. Equation 19.70 can be re-written as:

$$\ln P^{vap}(T)=A-\frac{B}{T} \tag{19.71}$$

with $B=\Delta H_{vap}/R$, and it is reasonably accurate for estimating the temperature dependence of the vapor pressure over small temperature ranges.

Correspondingly, the Antoine equation can be expressed by:

$$\ln P^{*}=A-\frac{B}{T+C} \tag{19.72}$$

or

$$\log P^{*}=\frac{\ln P^{*}}{\ln 10}=A-\frac{B}{T+C} \tag{19.73}$$

where A, B, and C are dimensional constants, P* is the vapor pressure and T is the temperature. Equation 19.73 is used to correlate vapor pressure accurately over the range from 1 to 200 kPa. The generalized Antoine equation is expressed as:

$$\ln P_r^* = A' - \frac{B'}{T_r + C'} \tag{19.74}$$

where A′, B′, and C′ are constants that depend upon the acentric factor. Tables 19.6 and 19.7 represent the generalized Antoine constants derived from the vapor pressure predictions of the RSK equation and the PR equation, respectively. Table 19.7 shows the transformation required to convert values of A′, B′, and C′ to Antoine constants of A, B, and C for use with various temperature and pressure units in Equation 19.71. An Excel spreadsheet program (Example 19.1.xlsx) has been developed to determine the vapor pressure, molar volumes, and fugacity coefficients from Soave–Redlich–Kwong (SRK) equation of state. The spreadsheet also compares the vapor pressure using the generalized Antoine's equation with the polynomial Equation 19.66, and the calculated vapor pressure from SRK equation. Another spreadsheet program (Example 19.2.xlsx) has been developed to compare the vapor pressure of components using the Antoine Equation 19.72 and coefficients in Table 19.8 with the Clausius–Clapeyron's Equation

Table 19.7 Transforming Antoine constants A´, B´ and C´ from Table 19.5 to dimensional constants A, B and C for use in Equation 19.71.

$A = A' + 1n\ P_c$	The units of P_c must be absolute pressure Vapor pressure P* in Equation 19.71 will have the same units as P_c
$B = B'\ T_c$	The units of T_c must be absolute temperature in K or °R
$C = C'\ T_c$	Temperature T in Equation 19.71 will have the same as T_c
$C = C'\ T_c + 273.15$	Temperature T in Equation 19.71 in °C and T_c in K
$C = C'\ T_c + 459.56$	Temperature T in Equation 19.71 in °F and T_c in °R

(Source: DeDoes, A. J., *et al.*, CEP, pp 39-44, Jan. 2007).

Table 19.8 Antoine coefficients for selected substances.

Substance	A	B	C	Temperature range, °C
Acetone	7.63132	1566.69	273.419	57 to 205
Water	8.01767	1715.70	234.268	100 to 265
Benzene	6.87987	1196.76	219.161	8 to 80
Toluene	6.95087	1342.31	219.187	-27 to 111
Ethylene glycol	8.09083	2088.936	203.454	50 to 150
Hexane	6.91058	1189.64	226.280	-30 to 170
p-Xylene	6.99053	1453.430	215.310	27 to 166
Ethanol	8.11220	1592.864	226.184	20 to 93
Acetic acid	8.02100	1936.010	258.451	18 to 118

(*Continued*)

Table 19.8 Antoine coefficients for selected substances. (*Continued*)

Substance	A	B	C	Temperature range, °C
Acetaldehyde	7.20812	1099.810	233.945	-82 to 20
Methanol	8.08097	1582.271	239.726	15 to 84
Methyl ethyl ketone	7.06356	1261.340	221.969	43 to 88
Chloroform	6.95465	1170.966	226.232	-10 to 60
Ethylenediamine	8.09831	1893.720	245.676	11 to 117
4-Methyl-2-pentanol	8.46706	2174.869	257.780	25 to 133
Dichloromethane	7.08030	1138.910	231.450	-44 to 59
1,3-Butadiene	6.85364	933.586	239.511	-75 to -2
Styrene	7.50233	1819.810	248.662	-7 to 145
Ethyl acetate	7.10179	1244.950	217.881	16 to 76
Vinylacetylene	7.02515	999.110	235.817	-93 to 5
Acetic anhydride	7.69301	1781.29	230.395	2 to 140
Dichlorosilane[§]	7.18600	1224.50	273.16	-45 to 121
Trichlosilane	6.95524	1102.900	238.865	-81 to 32
Silicon tetrachloride	7.02404	1212.890	235.910	-63 to 57
Hydrogen chloride[†]	7.44899	868.358	274.904	-85 to 36

(Source: Vapor-Liquid Equilibrium Data Collection of the DECHEMA Chemistry Data Series, except where noted. The form of the equation is $\log_{10} P^{vap} = A - B/(T+C)$ with pressure in mmHg and temperature in °C).

[§] Bawa, M. S., Texas Instruments (1988).

[†] Ohe, S., Computer-Aided Data Book of Vapor Pressure, Data Book Publishing Company, Tokyo, Japan (1976).

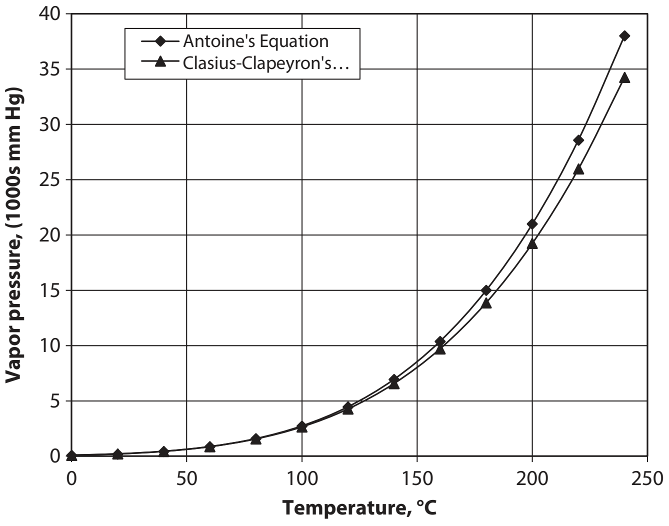


Figure 19.15 Vapor pressure of Antoine's vs. temperature.

19.69 over a range of temperature. The saturated vapor pressure curves of acetone predicted by Equations 19.69 and 19.72 are shown in Figure 19.15. The parameters taken from Reid *et al.* [21] and have the values T_{nb} = 329.4 K, $\Delta H_{vap} = 6960\,\text{cal/gmol}$, R = 1.98721 cal/gmol·K. At pressures < 10 atm., the Clausius–Clapeyron equation reproduces the experimental data very closely and would be appropriate for use in vapor–liquid equilibrium (VLE) calculations. At high pressure, the predictions are not so good, and by comparison the Antoine coefficients can only be obtained by regression of experimental vapor pressure data. If the constant ΔH_{vap} in the Clausius–Clapeyron equation is treated as an adjustable parameter and fitted to the experimental vapor pressure data, a value of 7175 cal/gmol produces a curve that fits the data accurately as the Antoine equation [7].

Example 19.1: Soave–Redlich–Kwong Prediction of the Vapor Pressure

For n-pentane, T_c = 469.76K, P_c = 33.7 bar, $\boldsymbol{\omega} = \mathbf{0.252}$. Use these data to verify that the SRK equation predicts a vapor pressure of 1.05 bar at 309.2 K. Also verify these data with the generalized Antoine equation. R = 83.14 cm^3-bar/mol-K.

Solution

The Excel spreadsheet Example 19.1.xls has been developed to calculate the molar volumes of liquid and vapor of any component such as pentane, its fugacity coefficient in liquid and vapor phases, and the vapor pressure of n-pentane using the SRK, the generalized Antoine equation and a polynomial Equation 19.66. The percentage deviation in the value of the vapor pressure between the SRK equation and the generalized Antoine's equation can be calculated. Equation 19.74 depends on the reduced temperature and acentric factor, and Table 19.5 provides the constants A′, B′, and C′ for SRK at a range of reduced temperature, T_r ($0.64 \le T_r \le 1.0$). The spreadsheet calculates the vapor pressure of SRK equation by using a Goal seek/Solver optimization tool in the Excel spreadsheet. The SRK (Equation 5) in Table 19.3 is:

$$P = \frac{RT}{V - b_{SRK}} - \frac{a_{SRK}(T)}{V(V + b_{SRK})} \tag{19.75}$$

where

$$a_{SRK}(T) = 0.42748R^2T_c^2\left[1 + f_\omega(1 - T_r^{0.5})\right]^2/P_c \tag{19.76}$$

$$b_{SRK} = 0.08664RT_c/P_c \tag{19.77}$$

$$f_\omega = 0.48 + 1.574\omega - 0.176\omega^2 \tag{19.78}$$

The constant parameter a_{SRK} in Equation 19.76 is calculated from Equation 19.78, with known and calculated values of the critical temperature, reduced temperature, critical pressure and the Universal gas constant; the constant parameter b_{SRK} is calculated from Equation 19.77. Equation 19.75 is rearranged in terms of the molar volume as a cubic equation and is solved for liquid and vapor molar volumes using the developed Excel spreadsheet (Example 19.1.xlsx). The SRK equation is now expressed in terms of the molar volume as:

$$f(V)=V^3-\frac{RT}{P}V^2+\left(\frac{a_{SRK}-b_{SRK}-Pb_{SRK}^2}{P}\right)V-\frac{a_{SRK}\,b_{SRK}}{P}=0 \tag{19.79}$$

Using the calculated vapor pressure from the generalized Antoine Equation 19.74, the molar volumes (three real roots) are determined. The liquid molar volume and the vapor molar volume (i.e., the lowest and highest values) are then chosen from these three roots, and are used to determine the fugacity of the liquid and vapor phases from the following equations [44]:

$$Z=\frac{PV}{RT} \tag{19.80}$$

$$h=\frac{b_{SRK}}{V} \tag{19.81}$$

$$\ln\phi=Z-1-\ln Z-\ln(1-h)-\frac{a_{SRK}}{b_{SRK}\,RT}\ln(1+h) \tag{19.82}$$

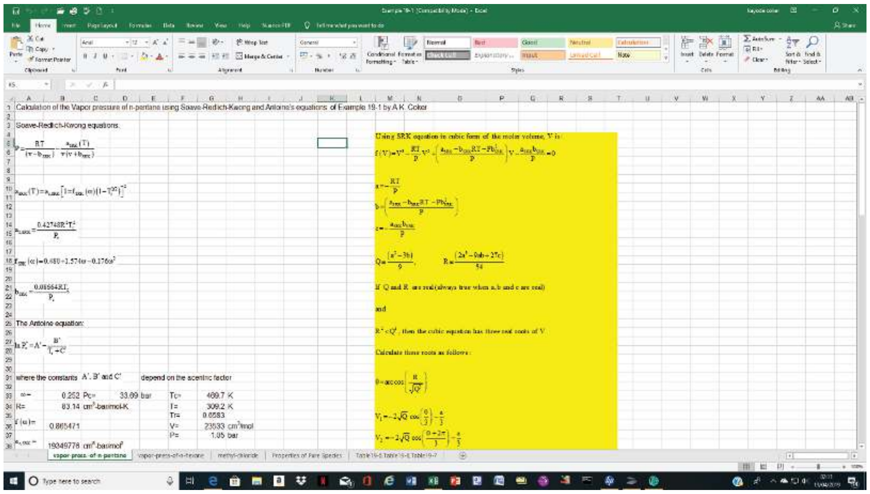
Equation 19.75 is rearranged in the form

$$f(P)=P-\frac{RT}{V-b_{SRK}}+\frac{a_{SRK}(T)}{V(V+b_{SRK})}=0 \tag{19.83}$$

Equation 19.83 in the Excel spreadsheet (Example 19.1.xls) is set to zero (cell B44) and is optimized by changing cell G37 using Solver or Go-Seek optimizer in the spreadsheet. This is carried out for known value of the calculated molar volume V, by changing the value of the vapor pressure in the cell. When the final value of Equation 19.83 becomes 0, then the computed value of P (cell G37) gives the required vapor pressure. Since the calculated fugacity coefficients for both liquid and vapor are equal, the vapor pressure as calculated from SRK equation is the estimated known vapor pressure at the specified temperature. In this example, the vapor pressure of n-pentane at 309.2 K is 1.05 bar. Figures 19.16a–c show snap shots of the Excel spreadsheet for calculating the vapor pressure using the SRK model.

Example 19.2: Use of the Clausius–Clapeyron Equation

The vapor pressure of liquid 2,2,4 trimethyl pentane at various temperatures is given in Table 19.9. Estimate the heat of vaporization of this compound at 25°C.



Calculation of the Vapor pressure of n-pentane using Soave-Redlich-Kwong and Antoine's equations of Example 19-1 by A.K. Coker

Soave-Redlich-Kwong equations:

$$P=\frac{RT}{(v-b_{SRK})}-\frac{a_{SRK}(T)}{v(v+b_{SRK})}$$

$$a_{SRK}(T)=a_{c,SRK}\left[1+f_{SRK}(\omega)(1-T_r^{1/2})\right]^2$$

$$a_{c,SRK}=\frac{0.42748R^2T_c^2}{P_c}$$

$$f_{SRK}(\omega)=0.480+1.574\omega-0.176\omega^2$$

$$b_{SRK}=\frac{0.08664RT_c}{P_c}$$

The Antoine equation:

$$\ln P_r^s=A'-\frac{B'}{T_r+C'}$$

where the constants A', B' and C' depend on the acentric factor

ω= 0.252 Pc= 33.69 bar Tc= 469.7 K
R= 83.14 cm³-bar/mol-K T= 309.2 K
Tr= 0.6583
f(ω)= 0.865471 V= 23533 cm³/mol
P= 1.05 bar
$a_{c,SRK}$ = 19349778 cm⁶-bar/mol²

Using SRK equation in cubic form of the molar volume, V is:

$$f(V)=V^3-\frac{RT}{P}V^2+\left(\frac{a_{SRK}-b_{SRK}RT-Pb_{SRK}^2}{P}\right)V-\frac{a_{SRK}b_{SRK}}{P}=0$$

$$a=-\frac{RT}{P}$$

$$b=\left(\frac{a_{SRK}-b_{SRK}RT-Pb_{SRK}^2}{P}\right)$$

$$c=-\frac{a_{SRK}b_{SRK}}{P}$$

$$Q=\frac{(a^2-3b)}{9},\qquad R=\frac{(2a^3-9ab+27c)}{54}$$

If Q and R are real (always true when a, b and c are real)

and

$R^2<Q^3$, then the cubic equation has three real roots of V

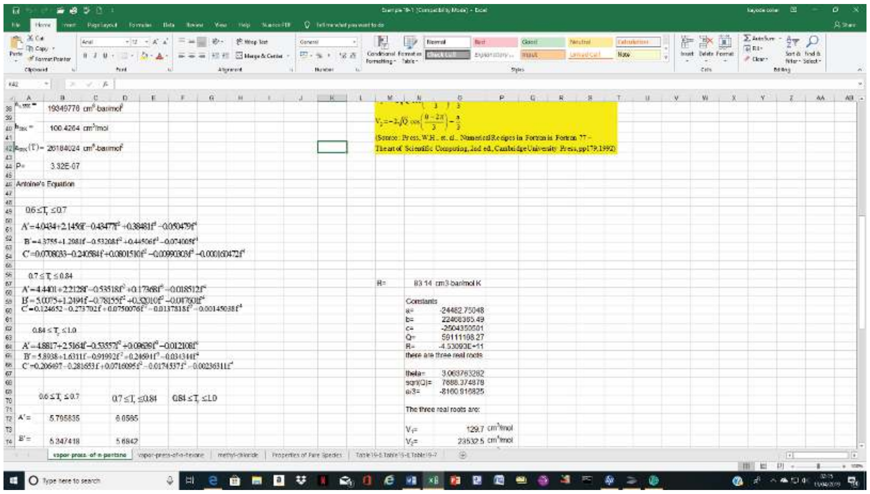
Calculate three roots as follows:

$$\theta=\arccos\left(\frac{R}{\sqrt{Q^3}}\right)$$

$$V_1=-2\sqrt{Q}\cos\left(\frac{\theta}{3}\right)-\frac{a}{3}$$

$$V_2=-2\sqrt{Q}\cos\left(\frac{\theta+2\pi}{3}\right)-\frac{a}{3}$$

(**a**)



$a_{c,SRK}$ = 19349778 cm⁶-bar/mol²
b_{SRK} = 100.4264 cm³/mol
$a_{SRK}(T)$ = 26184024 cm⁶-bar/mol²
P= 3.32E-07

Antoine's Equation

$0.6\le T_r\le 0.7$

$$A'=4.0434+2.1456f-0.43477f^2+0.38481f^3-0.050479f^4$$
$$B'=4.3755+1.2081f-0.53208f^2+0.44506f^3-0.074005f^4$$
$$C'=0.0708033-0.240584f+0.0801510f^2-0.0990303f^3-0.000160472f^4$$

$0.7\le T_r\le 0.84$

$$A'=4.4401+2.2128f-0.53518f^2+0.17368f^3-0.018512f^4$$
$$B'=5.0075+1.2494f-0.78155f^2+0.32010f^3-0.047601f^4$$
$$C'=0.124652-0.273702f+0.0750076f^2-0.0137818f^3-0.00145038f^4$$

$0.84\le T_r\le 1.0$

$$A'=4.8817+2.5164f-0.53557f^2+0.09639f^3-0.012108f^4$$
$$B'=5.8038+1.6311f-0.91992f^2+0.24691f^3-0.034344f^4$$
$$C'=0.206497-0.281653f+0.0716095f^2-0.0174537f^3-0.00236311f^4$$

	$0.6\le T_r\le 0.7$	$0.7\le T_r\le 0.84$	$0.84\le T_r\le 1.0$
A'=	5.795835	6.0585	
B'=	6.247418	5.6842	

$$V_3=-2\sqrt{Q}\cos\left(\frac{\theta-2\pi}{3}\right)-\frac{a}{3}$$

(Source: Press, W.H., et. al., Numerical Recipes in Fortran is Fortran 77 – The art of Scientific Computing, 2nd ed., Cambridge University Press, pp179, 1992)

R= 83.14 cm3-bar/mol K

Constants
a= -24482.75048
b= 22408365.49
c= -2504350501
Q= 59111108.27
R= -4.53093E+11
there are three real roots

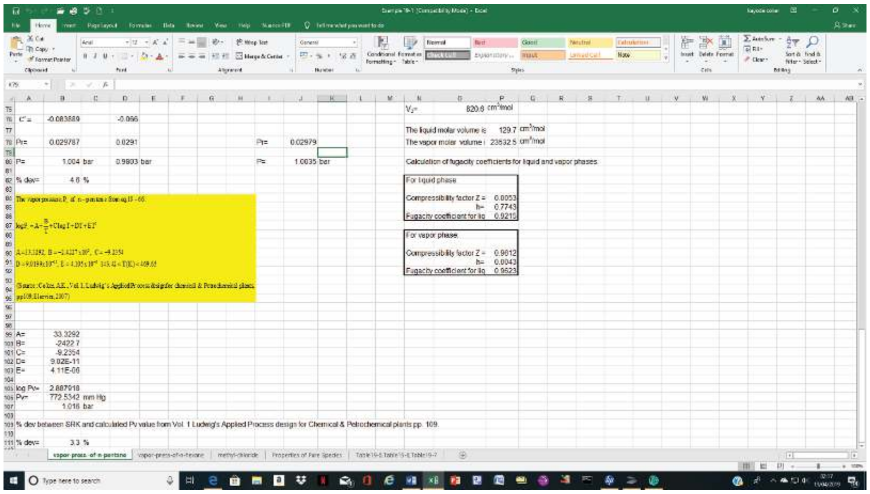
theta= 3.063763282
sqrt(Q)= 7688.374878
a/3= -8160.916825

The three real roots are:

V₁= 129.7 cm³/mol
V₂= 23532.5 cm³/mol

(**b**)

Figure 19.16a, b Snapshot of the Excel spreadsheet for calculating the vapor-pressure using SRK method and Antoine's Equation (Example 19.1).



C'=	-0.083889	-0.066		
Pr=	0.029787	0.0291	Pr=	0.02979
P=	1.004 bar	0.9803 bar	P=	1.0035 bar
% dev=	4.6 %			

The vapor pressure P_v of n-pentane from eq.15-66

$$\log P_v=A-\frac{B}{T}+C\log T+DT+ET^2$$

A=33.3292, B=−2.4227×10³, C=−9.2354
D=9.0199×10⁻¹¹, E=4.105×10⁻⁶ 143.42<T(K)<469.65

(Source: Coker, A.K., Vol 1, Ludwig's Applied Process design for chemical & Petrochemical plants, pp109, Elsevier, 2007)

A= 33.3292
B= -2422.7
C= -9.2354
D= 9.02E-11
E= 4.11E-06

log Pv= 2.887918
Pv= 772.5342 mm Hg
1.016 bar

% dev between SRK and calculated Pv value from Vol. 1 Ludwig's Applied Process design for Chemical & Petrochemical plants pp. 109.

% dev= 3.3 %

V₃= 820.6 cm³/mol

The liquid molar volume is 129.7 cm³/mol
The vapor molar volume i 23532.5 cm³/mol

Calculation of fugacity coefficients for liquid and vapor phases.

For liquid phase

Compressibility factor Z = 0.0053
h= 0.7743
Fugacity coefficient for liq 0.9215

For vapor phase

Compressibility factor Z = 0.9612
h= 0.0043
Fugacity coefficient for liq 0.9623

(**c**)

Figure 19.16c Snapshot of the Excel spreadsheet for calculating the vapor-pressure using SRK method and Antoine's Equation (Example 19.1).

Table 19.9 Vapor pressure of liquid 2,2,4 trimethyl pentane vs. temperature.

Vapor pressure (kPa)	0.667	1.333	2.666	5.333	8.000	13.33	26.66	53.33	101.32
Temperature (°C)	−15.0	−4.3	7.5	20.7	29.1	40.7	58.1	78.0	99.2

Solution

The Excel spreadsheet Example 19.2.xls has been developed to estimate the heat of vaporization and Figure 19.17 shows the trendline from the spreadsheet indicating that ΔH_{vap} is constant over the whole range of temperature. The value of the slope from the trendline is:

$$\frac{\Delta H_{vap}}{R} = 4.240 \times 10^3$$

R = 8.314 J/mol•K and the estimated $\Delta H_{vap} = 8.314 \times 4.24 \times 10^3$ J/mol

The calculated heat of vaporization, ΔH_{vap} = 35.16 kJ/mol.

Example 19.3: Construction of Temperature Composition Diagram and Equilibrium Diagram and Plot of Relative Volatility vs. Mole fraction of Phenol–Orthocresol Mixture

The following vapor pressures were obtained for phenol and orthocresol.

Temperature (K)	Vapor pressure (kN/m²)Othocresol	Phenol
387.0	7.70	10.0
387.9	7.94	10.4
388.7	8.21	10.8
389.6	8.50	11.2
390.3	8.76	11.6
391.1	9.06	12.0
391.9	9.40	12.4
392.7	9.73	12.9
393.3	10.00	13.3

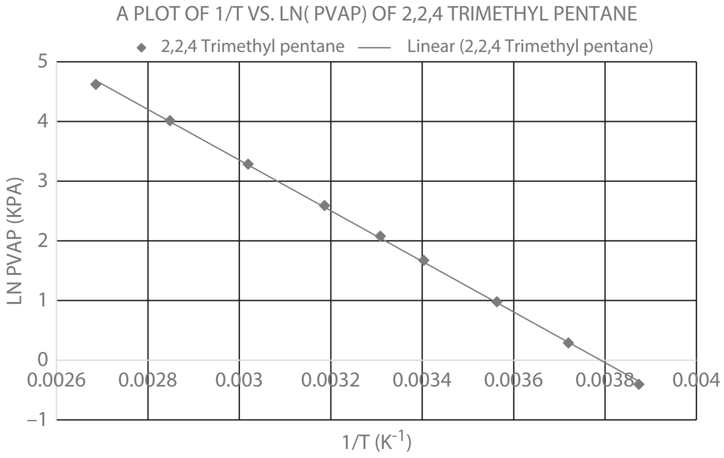


Figure 19.17 A plot of 1/T vs. ln P^{vap} of 2, 2, 4 trimethyl pentane.

Assuming Raoult's and Dalton's laws apply; construct the following data for a total pressure of 10 kN/m^2

1. A temperature composition diagram.
2. An x-y diagram.
3. Relative volatility against mole fraction of phenol in liquid.

Solution

If Raoult's law and Dalton's law apply, the following equations may be used.

$$\pi = p_A + p_B \tag{19.2}$$

$$p_A = y_A \cdot P = x_A \cdot p_A^o \tag{19.31}$$

where

π = the total pressure
p_A = the partial pressure of component A
p_B = the partial pressure of component B
p_A^o = the vapor pressure of component A.
p_B^o = the vapor pressure of component B.

Hence from Equations 19.2 and 19.30

$$x_A = \frac{\left(\pi - p_B^o\right)}{\left(p_A^o - p_B^o\right)} \tag{19.84}$$

$$y_A = \frac{x_A \cdot p_A^o}{\pi} \tag{19.85}$$

Table 19.10 Mole fractions x_A, y_A, and α_{A-B} in Example 19.3.

Temperature K	x_A **(from Eq. 19.84)**	y_A **(From Eq. 19.85)**	$\alpha_{A-B} = \frac{p_A^o}{p_B^o}$ **(From Eq. 19.86)**
387.0	1.0	1.0	1.300
387.9	0.837	0.871	1.310
388.7	0.691	0.746	1.315
389.6	0.556	0.622	1.318
390.3	0.437	0.506	1.324
391.1	0.320	0.384	1.325
391.9	0.200	0.248	1.319
392.7	0.085	0.110	1.326
393.3	0	0	1.330

The relative volatility α_{A-B} is defined in terms of the more volatile component (e.g., Phenol) by:

$$\alpha_{A-B} = \frac{p_A^o}{p_B^o} \tag{19.86}$$

Table 19.10 shows the results of the computation.

Figures 19.18 and 19.19 show the temperature composition diagram and the equilibrium diagram, respectively. The values obtained for the relative volatility vary between 1.300 and 1.330 and the average value of 1.319 could be used in subsequent computations. The Excel spreadsheet Example 19.3.xls shows the calculations of x_A, y_A, and α_{A-B}.

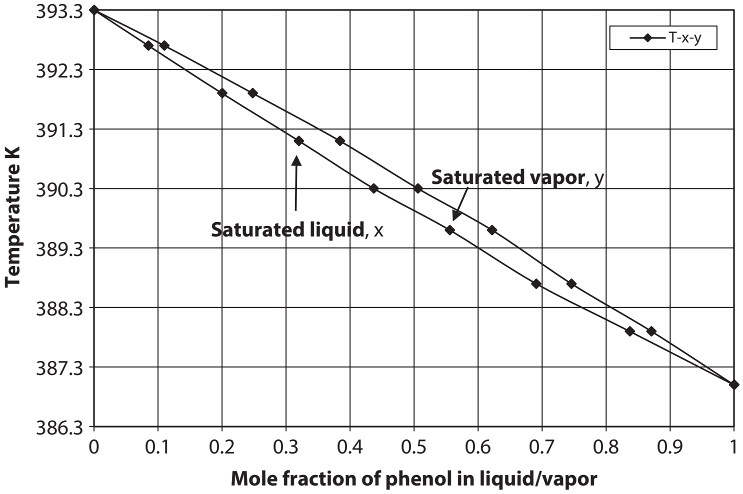


Figure 19.18 Phase diagram for the mixture phenol and orthocresol at 10 kN/m².

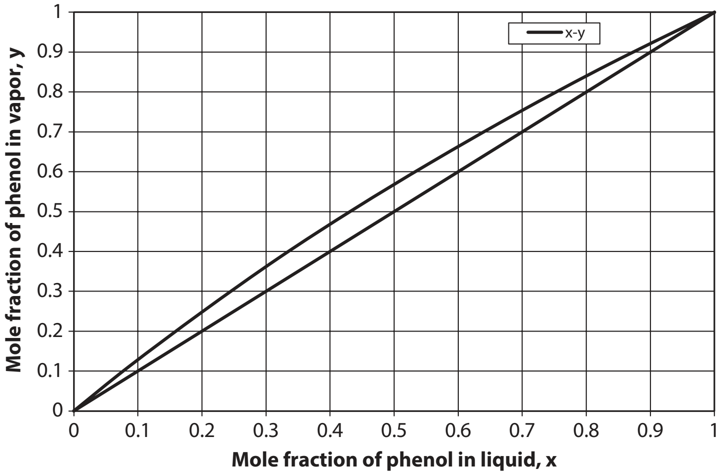


Figure 19.19 VLE data for the mixture phenol and orthocresol at 10 kN/m².

19.13 Azeotropic Mixtures

Azeotrope is derived from the ancient Greek words meaning "*to boil unchanged*", and is defined as a mixture of composition at which the equilibrium liquid and vapor compositions are equal. Figures 19.5, 19.6a, 19.6b, 19.6c, and 19.7 are phase diagrams for normal systems, showing that the concentration of the less volatile component increases with increase in the dew and bubble points. However, if the components exhibit strong physical and chemical interaction, the resulting phase diagrams are significantly different. In such systems, there is a critical composition (i.e., the point of intersection of the equilibrium curve with the 45° line) for which the vapor and liquid compositions are identical. At this point, the components cannot be separated at the given pressure, and such mixtures are known as azeotropes.

Azeotropes are common when polar organic compounds, such as alcohols, ketones, ethers and acids are mixed with water, and this complicates their separation by distillation. In practice, there will be no change in liquid and vapor compositions from tray to tray, and hence a simple distillation process will not separate such mixtures, it is useful to examine their deviations from Raoult's Law.

Azeotropes occur in various systems as binary, ternary and multicomponent mixtures, and can be "homogeneous" (single liquid phase) or "heterogeneous" (two liquid phases). They can exhibit a minimum-boiling point or maximum-boiling point.

A minimum-boiling azeotrope boils at a temperature lower than either of the pure components. Consequently, when distilling a binary system made up of these components, the top product is the azeotrope, and the bottom is one or the other of pure components depending on the "side" of the azeotrope that the feed was. A maximum-boiling azeotrope boils at a temperature higher than either of the pure components and leaves from a bottom of the column and the top product is the high boiling point component containing the more volatile component (MVC) at low concentrations. The top product is the MVC when it is present at high concentrations. When liquid phase separation occurs, the boiling temperature of the mixture as well as the vapor phase composition remains constant until one of the liquid phases disappears. Under such conditions and depending on the composition, a mixture of the two liquids will leave the top of the column while either of the components will leave at the bottom.

Azeotrope mixtures consist of two or more components, and are surprisingly common in distillation systems. It is therefore essential with any new system to determine whether an azeotrope exists. Fortunately, if experimental data are unavailable, reference can be made to excellent lists for known azeotropic systems, with vapor pressure information [45–47]. Typical forms of representation of azeotropic data are shown in Figures 19.20a and 19.20b and 19.21a and 19.21b, respectively. These are homogeneous, with one liquid phase at the azeotrope point. Figures 19.22a and 19.22b illustrate a heterogeneous azeotrope where two liquid phases are in equilibrium with one vapor phase. The system butanol–water is an example of the latter. Chloroform–methanol and acetone–chloroform are examples of binary homogeneous azeotropes with "minimal-boiling point" and "maximum-boiling point," respectively.

A "minimum" boiling binary azeotrope exhibits a constant composition as shown by its crossing of the $x = y$ along the 45° line in Figure 19.20b, which boils at a lower temperature than either of its pure components. This class of azeotrope results from positive deviations from Raoult's Law. Likewise, the "maximum" (Figure 19.21b) boiling binary azeotrope represents a negative deviation from Raoult's Law and exhibits a constant boiling point greater than either of the pure components. At the point where the equilibrium curve crosses $x = y$, 45° line, the composition is constant and cannot be further purified by normal distillation. Both minimum and maximum boiling point azeotropes can be modified by changing the system pressure and/or addition of a third component, which should be chosen such that will form a minimum boiling azeotrope with one of the original pair. To be effective the new azeotrope should boil well below or above the original azeotrope. Using this technique, one of the original components can often be recovered as a pure product, while still obtaining the second azeotrope for separate purification.

In a minimum-boiling azeotrope, the molecular species repel each other, whereas in a maximum-boiling azeotrope they attract each other. However, if the azeotrope occurs in the composition range in which the two liquids are immiscible, phase splitting occurs and a heterogeneous azeotrope is formed. Table 19.11 lists some examples of binary mixtures that form azeotropes.

For a "minimum" boiling azeotrope the partial pressures of the components will be greater than predicted by Raoult's Law, and the activity coefficients will be >1.0.

$$\gamma = (y_i P_{Tot})/(x_i p_i^*) \tag{19.87}$$

where

p_i^* = vapor pressure of component i, at temperature
$P_{Tot} = \pi$ = total pressure
$\gamma = a_i/x_i$ = activity coefficient of component, i
p_i = partial pressure of component i.

Raoult's Law: $p_i = x_i P_i^* = y_1 P_{Tot}$

For "maximum" boiling azeotropes, the partial pressures will be less than predicted by Raoult's Law and the activity coefficients will be less than 1.0.

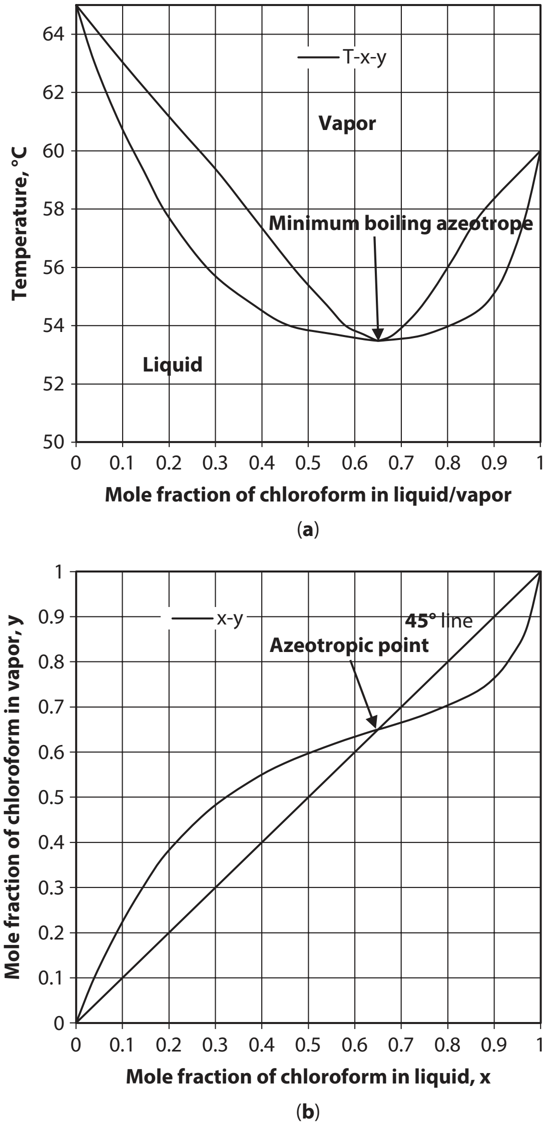


Figure 19.20 (a) Phase diagram for the mixture of chloroform and methanol at 1 atm, (b) VLE diagram for chloroform and methanol mixture at 1 atm.

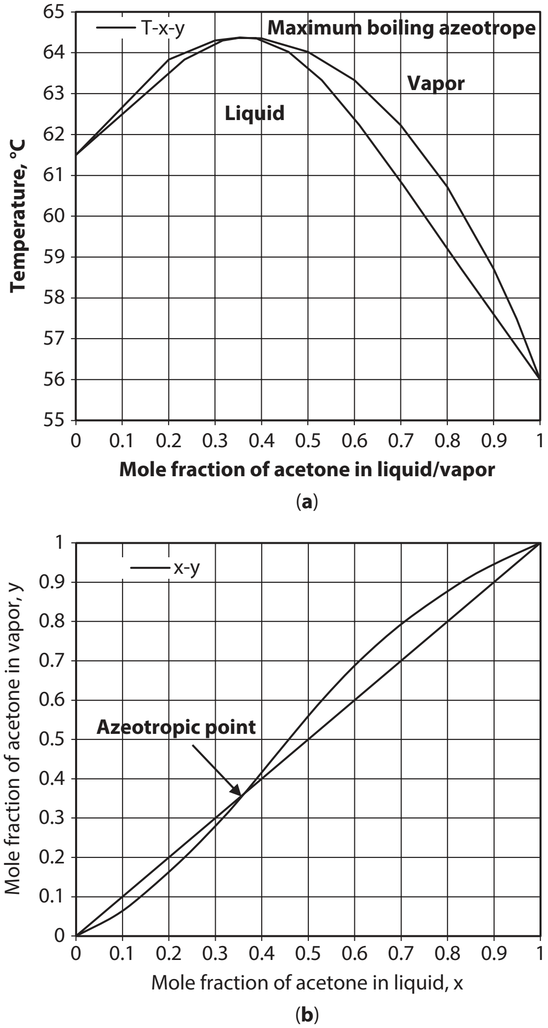
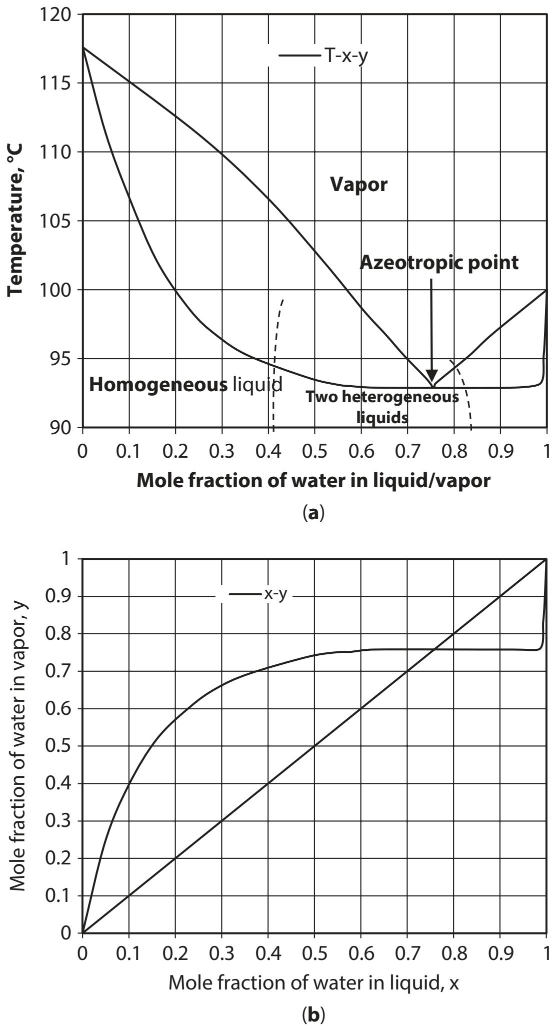


Figure 19.21 (a) Phase diagram for the mixture chloroform and acetone at 1 atm, (b) Vapor–liquid equilibrium diagram for chloroform acetone mixture at 1 atm.

In reference to distillation conditions, the azeotrope represents a point in the system where the relative volatilities reverse. This applies to either type of azeotrope but the direction of reversal is opposite. For example, in Figure 19.20b the lower portion of the x-y diagram shows that $y_i > x_i$, while at the upper part, the $y_i < x_i$. In any distillation, without addition of an azeotrope "breaker" or solvent to change the system characteristics, if a feed of composition 30% x_i were used, the column could only yield a bottom product, which approaches pure x_2 and an azeotrope composition of about 65% x_i and 35% x_2 at the top. The situation would be changed only to the extent of recognizing that if the feed came in above the azeotropic point, the bottoms product would be the azeotrope composition.

Successful azeotropic distillation is largely dependent on the correct choice of entrainer (i.e., solvent). Entrainers have been selected on the basis of making a separation distillation possible. Stichlmair and Fair [48] have presented criteria for entrainer selection, which are illustrated in Table 19.12 and Figure 19.23. The entrainer, selected such that the constituents *a* and *b* of the azeotropic mixture to be separated are both origins or termini of distillation lines, and such systems always have distillation borders. The entrainer must also permit the recovery of products *a* and *b* at the top or bottom of distillation columns.



(**a**)

(**b**)

Figure 19.22 (a) Phase diagram for the mixture of water and 1-butanol at 1 atm, (b) VLE diagram for the mixture of water and 1-butanol at 1 atm.

Stichlmair and Fair [48] have presented a generalized process for separating out the higher boiling components, *b*. In this instance, the sequence of the separations differs from that in Figure 19.23. Both modifications of the generalized process are characterized by two separations performed on the convex side and one separation performed on the concave side of the border distillation line. Using the criteria selection of Table 19.12, a minimum azeotrope can be separated by an intermediate boiling or a high boiling entrainer instead of a low boiler, if additional azeotropes are formed. Smith [49] discusses azeotropic distillation in detail; Widalgo and Seider [50] have presented an excellent review of azeotropic distillation. References [51]–[57] describe azeotropic design techniques.

Generally, all activity coefficient models are capable of representing phase diagrams of the type shown in Figures 19.10c and 19.10d, but not always to same level of accuracy. It is essential therefore, to plot the x-y diagram for each binary pair in the (multicomponent) mixture before using the VLE model in the design calculations. If the predicted diagrams do not provide satisfactory agreement with experimental data, other models should be considered. Otherwise, a poor designs result.

Commencing with the equality of the fugacities of the species in the vapor and liquid mixtures in Equation 19.12:

$$f_i^V = f_i^L, i = 1, \ldots\ldots, N \tag{19.12}$$

Table 19.11 Examples of binary mixtures with an azeotrope.

						Azeotropic data		
Formula	**Name**	**B.P., °C**	**Formula**	**Name**	**B.P., °C**	**xa, mol %**	**B.P., °C**	**Type**
H_2O	water	100	HBr	hydrobromic acid (a)	-67	16.7	126	max.
			HCl	hydrochloric acid (a)	-85	11.1	108.6	max.
			HNO_3	nitric acid (a)	86	38.3	121	max.
			$CHCl_3$	chloroform (a)	61.2	83.9	56.12	hetero
			CH_2Cl_2	dichloromethane (a)	40.1	93.3	38.1	hetero
			CH_3NO_2	nitromethane (b)	101.2	51.1	83.6	hetero
			C_2H_6O	ethanol (a)	78.3	90.4	78.2	min.
			$C_3H_6O_2$	methyl acetate (a)	57	87.2	56.4	min.
			C_3H_8O	n-propyl alcohol (a)	97.3	42.2	87.8	min.
			$C_4H_6O_2$	methyl acrylate (a)	80	72.9	71	min.
			C_4H_8O	tetrahydrofurane (a)	66	81.7	64	min.
			C_4H_8O	2-butanone (a)	79.6	66.2	73.4	min.
			$C_4H_8O_2$	butyraldehyde (a)	74	79.6	68	min.
			$C_4H_8O_2$	dioxane (b)	101.3	51.7	87.8	min.
			$C_4H_8O_2$	ethyl acetate (a)	77.1	68.2	70.4	hetero
			$C_4H_{10}O$	butyl alcohol (b)	117.4	75.2	92.7	hetero
			$C_4H_{10}O_2$	2-ethoxyethanol (b)	135.1	92.5	99.4	min.
			C_5H_5N	pyridine (b)	115	76.8	94	min.
			$C_5H_{12}O$	ethyl propyl ether (a)	63.6	83.1	59.5	hetero
			C_6H_6	benzene (a)	80.2	70.4	69.3	hetero
			C_7H_8	toluene (b)	110.7	55.6	84.1	hetero
$CHCl_3$	chloroform	61.2	CH_4O	methanol (b)	64.7	65.0	53.4	min.
			C_2H_6O	ethanol (b)	78.3	83.7	59.4	min.
			C_3H_6O	acetone (a)	56.1	36.7	64.4	max.
CH_4O	methanol	64.7	CH_2Cl_2	dichloromethane (a)	40.1	70.5	37.8	min.
			C_3H_6O	acetone (a)	56.2	80.2	55.5	min.
			C_4H_8O	2-butanone (b)	79.6	84	63.5	min.
			$C_4H_8O_2$	ethyl acetate (b)	77.1	69.9	62.3	min.
			C_6H_6	benzene (b)	80.2	61	57.5	min.
			C_7H_8	toluene (b)	110.7	86.5	63.8	min.
			C_7H_{16}	n-heptane (b)	98.5	76.5	59.1	min.
			C_8H_{18}	n-octane (b)	25.6	90.2	63	min.
C_2H_6O	ethanol	78.3	$C_4H_6O_2$	methylacrylate (b)	80	57.5	73.5	min.
			C_4H_8O	2-butanone (b)	79.6	50.0	74	min.
			$C_4H_8O_2$	ethylacetate (b)	77	53.8	71.8	min.
			$C_4H_{10}O_2$	Eth.methoxymeth (a)	65.9	77	64	min.
			$C_5H_{12}O$	ethylpropylether (a)	63.6	61	61.2	min.
			C_6H_{14}	n-hexane (a)	68.7	66.5	58.4	min.

(*Continued*)

Table 19.11 Examples of binary mixtures with an azeotrope. (***Continued***)

						Azeotropic data		
Formula	**Name**	**B.P., °C**	**Formula**	**Name**	**B.P., °C**	**xa, mol %**	**B.P., °C**	**Type**
			C_7H_{18}	toluene (b)	110.7	81	76.7	min.
C_3H_6O	acetone	56.2	C_6H_{12}	cyclohexane (b)	80.8	74.6	53	min.
			C_6H_{14}	hexane (b)	68.9	68.1	49.8	min.
			C_7H_{16}	heptane (b)	98.4	93.6	55.9	min.
$C_4H_{10}O_2$	2-ethoxyethanol	135.1	C_8H_{10}	ethylbenzene (b)	136.2	56.9	127.1	min.
			C_8H_{18}	octane (a)	125.8	61	116.1	min.
C_5H_5N	pyridine	115	C_7H_8	toluene (b)	110.7	75.3	110.1	min.
			C_8H_8	n-octane (b)	125.8	<92.9	128.8	min.
C_6H_6	benzene	80.2	C_3H_8O	2-propanol (b)	82.5	61	71.8	min.
			C_6H_{12}	cyclohexane (b)	80.8	53	77.6	min.

B.P. = boiling point, xa = mol % of component a (low boiler) at the azeotrope, p = 1.013 bar.
(Source: Stichlmair, J. G., and James R. Fair, Distillation: Principles and Practice, Wiley-VCH, 1998)

and substituting the expressions for the mixtures fugacities in terms of the mole fractions, activity coefficients, and fugacity coefficients in Equations 19.13 and 19.14:

$$y_i \phi_i^V P_{Tot} = x_i \gamma_i^L f_i^L, i = 1, \ldots.., N \tag{19.88}$$

Consider an ideal binary mixture $\left(\gamma_i^L = 1\right)$, at low pressure $\left(\phi_i^V = 1 \text{ and } f_i^L = P_i^s\right)$, where P_i^s is the vapor pressure of species i. Substituting in Equation 10.88,

Table 19.12 Criteria for entrainer selection for processes with distillation borders, see process in Figure 19.23.

Entrainer for the separation of a mixture with a minimum azeotrope
• Low boiler (lower than the minimum azeotrope)
• Medium boiler, which forms a new minimum azeotrope with the low boiling constituent of the given mixture.
• High boiler, which forms new minimum azeotropes with both constituents of the given mixture. At least one of them has to boil lower than the azeotrope of the given mixture.
Entrainer for the separation of a mixture with a maximum azeotrope
• High boiler (higher than the maximum azeotrope).
• Medium boiler, which forms a new maximum azeotrope with the high boiling constituent of the given mixture.
• Low boiler, which forms new maximum azeotropes with both constituents of the given mixture. At least one of them has to boil higher than the azeotrope of the given mixture.

(Source: Stichlmair, J.G., and J. R. Fair, Distillation: Principles and Practice, Wiley-VCH, 1998).

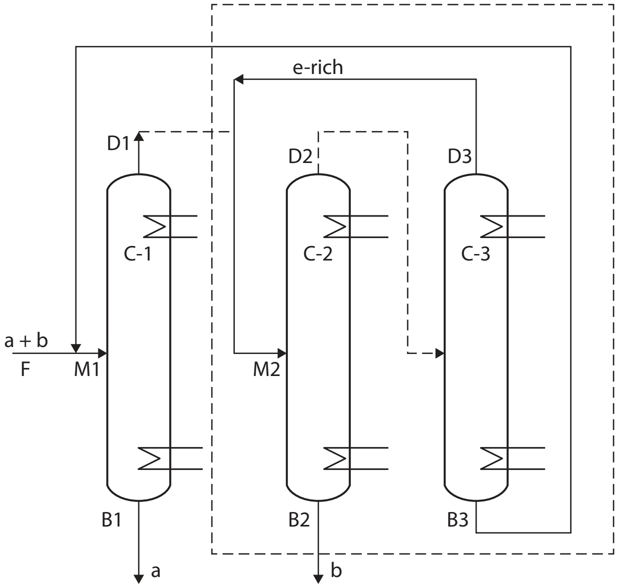


Figure 19.23 Process flow diagram for the separation of a binary mixture with a minimum azeotrope by using a low boiling entrainer. The feed is rich in a. Components a and b are obtained as pure bottoms products from columns C-1 and C-2, respectively.

$$y_1 P_{Tot} = x_1 P_1^s \tag{19.89}$$

and

$$y_2 P_{Tot} = x_2 P_2^s \tag{19.90}$$

Adding Equations 19.89 and 19.90:

$$P_{Tot} = x_1 P_1^s + x_2 P_2^s = x_1 P_1^s + (1 - x_1) P_2^s \tag{19.91}$$

$$= P_2^s + \left(P_1^s - P_2^s\right) x_1 \tag{19.92}$$

Equation 19.92 shows the relationship between the total pressure, P_{Tot} and the mole fraction of the more volatile species, which is characteristic of Raoult's law as shown in Figure 19.24 for benzene–toluene mixture. However, the mixture exhibits a positive deviation from ideality $\left(\gamma_i^L > 1, i = 1, 2\right)$ or Raoult's law, Equation 19.92 becomes

$$P_{Tot} = x_1 \gamma_1^L P_1^s + (1 - x_1) \gamma_2^L P_2^s \tag{19.93}$$

Therefore, if γ_1^L and γ_2^L are ≥ 1 for all compositions under isothermal boiling, then it is common for the bubble and dew point curves to reach a maximum at the same composition, that is the azeotropic point. This situation is illustrated in Figures 19.25a and 19.25b for the ethyl acetate-ethanol mixture. Figure 19.25a shows the bubble and dew point curves on a T-x-y diagram at 1 atmosphere (101.3 kPa). Note the minimum boiling azeotrope at 71.8°C, where $x_1 = y_1 = 0.54$. Feed streams having lower mole fractions cannot be purified beyond 0.54 in a distillation

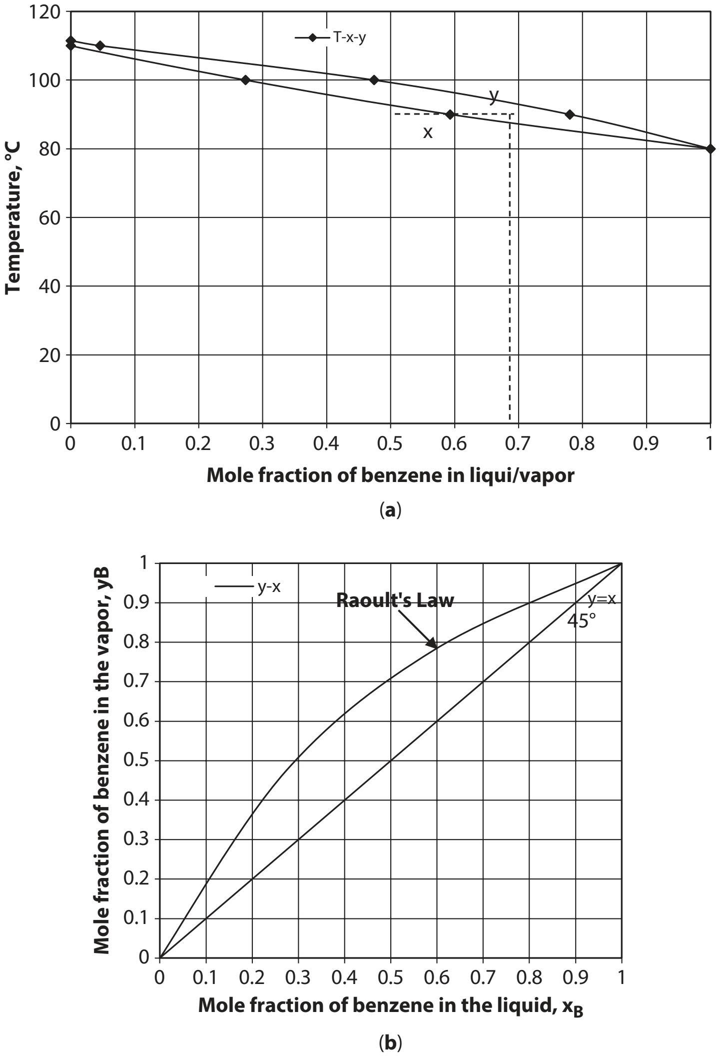


Figure 19.24 (a) Boiling point diagram for benzene–toluene mixture, total pressure = 760 mm Hg, (b) VLE diagram for the mixture benzene and toluene at 760 mm Hg total pressure.

column, and streams having higher mole fractions produce distillate mole fractions that are bounded by 0.54 and 1.0. Consequently, the azeotropic composition is commonly referred to as a distillation boundary.

Similarly, when the mixture exhibits a negative deviation from ideality or Raoult's law γ_1^L and γ_2^L are ≤ 1, both the bubble and dew point curves drop below the straight line that represents the bubble points for an ideal mixture. Figures 19.26a and 19.26b, respectively, show the phase and VLE diagrams of a water-formic acid binary mixture at 1 atmosphere (101.3 kPa). The T-x-y diagram shows the boiling point curving upward above ideality, and consequently the system is said to have the maximum boiling azeotrope at 107.6°C, where $x_1 = y_1 = 0.41$. In this instance, feed streams having lower mole fractions cannot be purified beyond 0.41 water in the bottoms product of a column, and feed streams having higher mole fractions have a lower bound of 0.41 water in the mole fraction of the bottoms product. The azeotropes in Figures 19.25a and 19.25b occur where $(\partial T/\partial x_1)_P = (\partial T/\partial y_1)_P = 0$, or equivalently, where $x_1 = y_1$. Therefore, azeotropic mixtures produce x-y curves that cross 45° line at the azeotropic composition. The slope of the curve at the crossing point is <1.0 for minimum-boiling azeotropes and >1.0 for maximum-boiling azeotropes, as shown in Figures 19.25a and 19.25b, respectively.

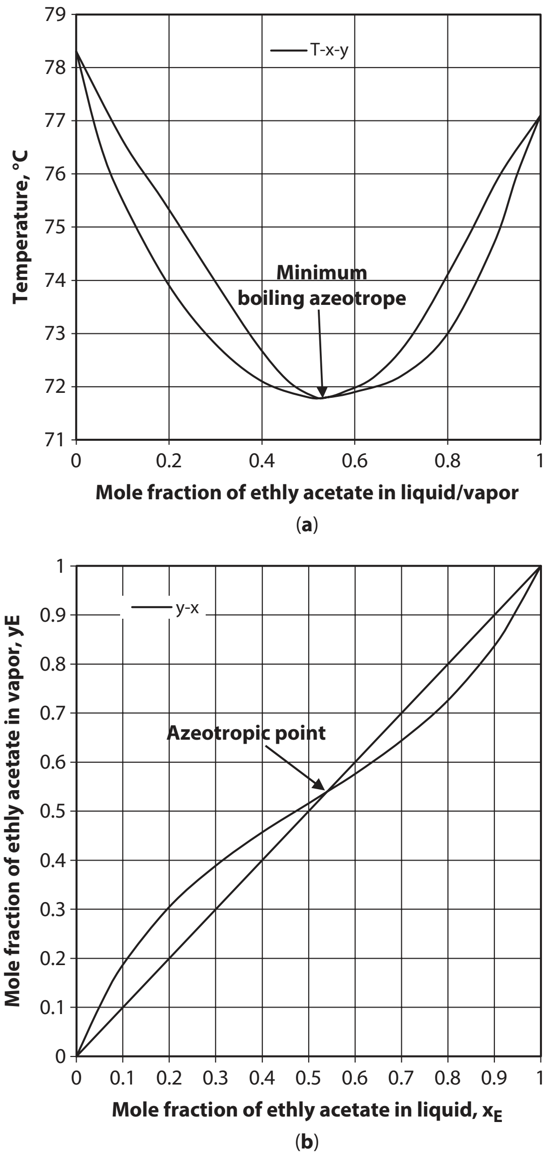
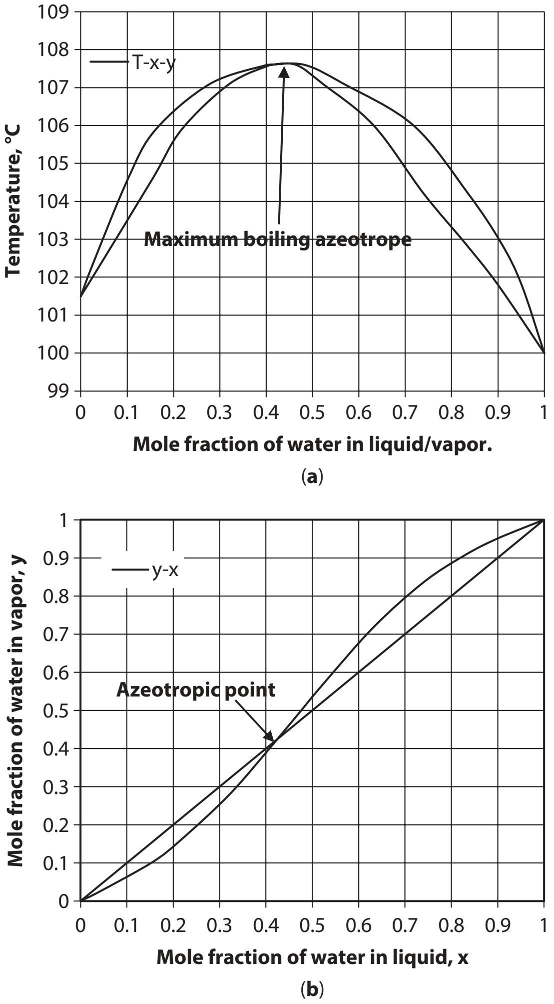


Figure 19.25 (a) Phase diagram for the mixture of ethyl acetate and ethanol at 1 atm, (b) VLE diagram for the mixture ethyl acetate and ethanol at 1 atm.

At homogeneous azeotrope, $x_i = y_i$, i = 1, ….C, the equilibrium constant, K_i for species becomes unity. The equilibrium constant for species i in Equation 19.27 becomes

$$K_i \equiv \frac{y_i}{x_i} = \frac{\gamma_i^L f_i^L}{\phi_i^V P_{Tot}} = 1, i = 1, \ldots\ldots N \tag{19.94}$$

where the degree of non-ideality is expressed by the deviations from unity of the activity coefficients, γ_i^L, for the liquid phase and the fugacity coefficients, ϕ_i^V for the vapor phase. At low pressure, $\phi_i^V = 1$ and $f_i^L = P_i^s$. Equation 19.94 reduces to



(a)

(b)

Figure 19.26 (a) Phase diagram for the mixture of water and formic acid at 101.3 kPa, (b) VLE diagram for the mixture of water and formic acid at 101.3 kPa.

$$K_i \equiv \frac{y_i}{x_i} = \gamma_i^L \frac{f_i^L}{P_{Tot}} = 1, i = 1, \ldots..N \tag{19.95}$$

Since the equilibrium constants for all of the species are unity at an azeotrope point, a simple distillation approaches this point, after which no further separation can occur. For this reason, an azeotrope is often referred to as a stationary or fixed pinched point. For a minimum-boiling azeotrope where the deviations from Raoult's law are sufficiently large $\left(\gamma_i^L >> 1.0\right)$, phase splitting may occur and a minimum-boiling heterogeneous azeotrope may occur that has a vapor phase in equilibrium with two liquid phases. A heterogeneous azeotrope occurs when the vapor–liquid envelope overlaps with the liquid–liquid envelope as shown in Figure 19.14. The author has developed phase diagrams of many binary systems from available literature using Microsoft Excel.

Example 19.4: Raoult's Law, Total Pressure Calculations for Multicomponent Mixture

A hydrocarbon liquid is a mixture at 55°F of the following components:

Component	Mole percent (%)
Iso-butane ($i - C_4H_{10}$)	41.5
Pentane (C_5H_{12})	46.5
n-hexane ($n - C_6H_{14}$)	12.0
	100.0

A vaporizer is to heat the mixture to 190°F at 110 psia.

Data from vapor pressure charts such as [27]:

Component	Vapor pressure, psia at 190°F
Iso-butane ($i - C_4H_{10}$)	235.0
Pentane (C_5H_{12})	65
n-hexane ($n - C_6H_{14}$)	26.0

Specific gravity of pure liquid at 55°F [28]:

Component	Specific gravity, SpGr.
Iso-butane ($i - C_4H_{10}$)	0.575
Pentane (C_5H_{12})	0.638
n-hexane ($n - C_6H_{14}$)	0.678

Moles in original liquid. Basis 100 gallons liquid

Assume Raoult's Law:

Component	Mols of component	
iso-butane ($i - C_4H_{10}$)	= 41.5 (8.33 × 0.575)/MW = 198.77/58.12 =	3.42
Pentane (C_5H_{12})	= 46.5 (8.33 × 0.638)/MW = 247.12/72.146 =	3.425
n-hexane ($n - C_6H_{14}$)	= 12 (8.33 × 0.678)/MW = 67.77/86.172 =	0.786
Total		7.631

Component	Mole fraction in the liquid phase	
iso-butane ($i - C_4H_{10}$)	x_1 = 3.42/7.631	0.448
Pentane (C_5H_{12})	x_2 = 3.425/7.63	0.449
n-hexane ($n - C_6H_{14}$)	x_3 = 0.786/7.631	0.103
Total		1.000

Mol fraction in vapor phase at 190°F. Raoult's Law:

$$y_i = p_i/\pi = (p_i^* x_i)/\pi \text{(for a binary system)}$$

$$y_i = (x_iP_i)/(x_1P_1 + x_2P_2 + x_3P_3) \text{ (for multicomponent mixtures)} \tag{19.96}$$

$$y_i = 0.448\ (235)/[(0.449)\ (65) + (0.448)\ (235) + (0.103)\ (26)]$$

$$= 105.28/[29.185 + 105.28 + 2.678] = 105.28/[137.143]$$

$$= 0.767$$

$$y_2 = 0.449\ (65)/137.143 = 0.212$$

$$y_3 = 0.103\ (26)/137.143 = 0.0195$$

$$\Sigma y_i = 0.998 \text{ (not rounded)}$$

Because, $P_{total} = (0.448)\ (235) + 0.449\ (65) + 0.103\ (26) = 137.14$ psia

This is greater than the selected pressure of 110 psia, therefore, for a binary mixture the results can be obtained without a trial-and-error solution. But, for the case of mixtures of three or more components, the trial-and-error assumption of the temperature for the vapor pressure will require a new temperature, re-determination of the component's vapor pressure, and repetition of the process until a closer match with the pressure is obtained.

Example 19.5: Determination of the Composition at an Azeotropic Temperature

The ethyl acetate (1)–ethyl alcohol (2) azeotrope has a boiling point of 71.8°C at a pressure of 101.325 kN/m^2 and a composition of 53.9 mole % ethyl acetate.

- **a.** Determine the composition of the azeotrope at 95°C from the table below:
- **b.** Comment briefly upon the implications of changing the pressure of an azeotrope distillation operation.

	Vapor Pressure	**kN/m²**
Temperature, °C	**Ethyl Acetate**	**Ethyl Alcohol**
70	79.03	72.55
80	110.75	108.32
90	151.38	158.27
100	195.11	217.49

The van Laar equations are:

$$\ln\gamma_1 = \frac{A}{\left[1+\left(\frac{Ax_1}{Bx_2}\right)\right]^2} \text{ and } \ln\gamma_2 = \frac{B}{\left[1+\left(\frac{Bx_2}{Ax_1}\right)\right]^2} \tag{19.97}$$

or by rearrangement,

$$A = \ln\gamma_1\left[1+\frac{x_2\ln\gamma_2}{x_1\ln\gamma_1}\right]^2 \text{ and } B = \ln\gamma_2\left[1+\frac{x_1\ln\gamma_1}{x_2\ln\gamma_2}\right]^2 \tag{19.98}$$

Solution

(a) The mole fraction in the liquid phase of ethyl acetate = x_1
The mole fraction in the vapor phase of ethyl acetate = y_1
The mole fraction in the liquid phase of ethyl alcohol = x_2
The mole fraction in the vapor phase of ethyl alcohol = y_2

At equilibrium and azeotropic conditions the following relationships hold:

$$f_i^V = f_i^L \text{ and } x_1 = y_1 \tag{19.12}$$

where

f_i^V = fugacity of component i in the vapor phase
f_i^L = fugacity of component i in the liquid phase

For a perfect gas and non-ideal solution

In the vapor phase, the fugacity:

$$f_1^V = y_1 \cdot \pi \tag{19.99}$$

where

π = the total pressure

For non-ideal liquid solution:

In the liquid phase, the fugacity:

$$f_1^L = \gamma_1 x_1 p_1^o \tag{19.100}$$

where

γ_1 = activity coefficient of ethyl acetate. It is a measure of the deviation from Raoult's law. It is greater than unity for positive deviations and less than unity for negative deviation.
p_1^o = vapor pressure of ethyl acetate.

Activity coefficients are defined for each component of a mixture as:

$$y_1\pi = \gamma_1 p_1^o x_1 \tag{19.101}$$

and

$$y_2\pi = \gamma_2 p_2^o x_2 \tag{19.102}$$

At azeotropic composition the values $x_1 = y_1$ and $x_2 = y_2$; so, the activity coefficients become:

$$\gamma_1 = \frac{\pi}{p_1^o} \tag{19.103}$$

and

$$\gamma_2 = \frac{\pi}{p_2^o} \tag{19.104}$$

Dividing Equation 19.103 by Equation 19.104 gives:

$$\frac{\gamma_1}{\gamma_2} = \frac{\pi/p_1^o}{\pi/p_2^o} = \frac{p_2^o}{p_1^o} \tag{19.105}$$

At the boiling point of 71.8°C, the vapor pressures of ethyl acetate, $p^o_{1,71.8°C}$ and ethyl alcohol, $p^o_{2,71.8°C}$ are by interpolation:

Ethyl acetate component	**Ethyl alcohol component**
$\frac{71.8-80}{70-80} = \frac{p^o_{1,71.8°C}-110.75}{79.03-110.75}$ $p^o_{1,71.8°C} = 84.74\,kN/m^2$ $= 85\ kN/m^2$	$\frac{71.8-80}{70-80} = \frac{p^o_{2,71.8°C}-108.32}{72.55-108.32}$ $p^o_{2,71.8°C} = 78.98\,kN/m^2$ $= 79\ kN/m^2$

The activity coefficients at the boiling temperature of 71.8°C are:

$$\gamma_1 = \frac{\pi}{p_1^o} = \frac{101.325}{85} = 1.192$$

and

$$\gamma_2 = \frac{\pi}{p_2^o} = \frac{101.325}{79} = 1.2826$$

The level of ethyl acetate in the mixture is 53.9 mole % and that of ethyl alcohol is 46.1 mole %.

The van Laar coefficients A and B are:

$$A = \ln(1.192)\left[1 + \frac{0.461\ln(1.2826)}{0.539\ln(1.192)}\right]^2$$
$$= 0.859$$

$$B = \ln(1.2826)\left[1 + \frac{0.539\ln(1.192)}{0.461\ln(1.2826)}\right]^2$$

$$= 0.826$$

Using the values of A and B, the activity coefficients γ_1 and γ_2 are computed for various values of x_1 and x_2 from the equations as shown below. The values of γ_1 and γ_2 are shown in the subsequent table.

$$\ln\gamma_1 = \frac{A}{\left[1 + \left(\frac{Ax_1}{Bx_2}\right)\right]^2} \text{ and } \ln\gamma_2 = \frac{B}{\left[1 + \left(\frac{Bx_2}{Ax_1}\right)\right]^2}$$

x_1	x_2	γ_1	γ_2	γ_1/γ_2
0.4	0.6	1.3505	1.1484	1.176
0.6	0.4	1.1407	1.3593	0.839
0.8	0.2	1.0329	1.7127	0.603

At the azeotrope temperature of 95°C, the values of p_1^o and p_2^o are determined by interpolation as follows:

$$\frac{95-100}{90-100} = \frac{p^o_{1,95°C} - 195.11}{151.38 - 195.11} \text{ and } \frac{95-100}{90-100} = \frac{p^o_{2,95°C} - 217.49}{158.27 - 217.49}$$

$$p^o_{1,95°C} = 173.245\text{kN/m}^2 \text{ and } p^o_{2,95°C} = 187.88\text{kN/m}^2$$

$$\frac{\gamma_1}{\gamma_2} = \frac{p_2^o}{p_1^o} = \frac{187.88}{173.245} = 1.0845$$

The corresponding liquid phase mole fraction x_1 is calculated by interpolation as follows:

$$\frac{x_1 - 0.6}{0.4 - 0.6} = \frac{1.0845 - 0.839}{1.176 - 0.839}$$

At 95°C, $x_1 = 0.454$

(b) Changing the pressure infers (by Charles' Law) that the temperature and therefore the boiling point will change directly. The composition will also change, so that an azeotrope may be avoided.

19.14 Bubble Point of Liquid Mixture

An equilibrium between a liquid and a vapor implies that the liquid is on the point of boiling, is producing bubbles of vapor within it, and the vapor is on the point of condensing or forming a dew. These conditions are referred to as saturated liquid and saturated vapor. At the prevailing total pressure, the liquid is at its bubble point temperature and the vapor at its dew point temperature. The amount of the vapor formed is assumed to be so small as to have no effect on the liquid composition. For a binary system,

$$x_1 = \frac{1-K_2}{K_1-K_2} \tag{19.106}$$

and

$$x_2 = 1.0 - x_1 \tag{19.107}$$

In the majority of cases involving multicomponent mixtures, no direct solution to Equation 19.106 exists. The basic definition of the equilibrium coefficient is

$$y_i = K_i x_i \tag{19.108}$$

For a stable system, the total mole fraction is unity, i.e.,

$$\sum y_i = \sum K_i x_i = 1.0 \tag{19.109}$$

Equation 19.109 is used to determine the bubble point temperature and pressure. In using it, the temperature or pressure is fixed, while the other parameter is varied until the criterion for a stable system is identified. A combination of temperature and pressure is altered, if the summation of the calculated vapor composition differs from unity. There is no known direct method that will allow a reasonable estimation of the amount of change required. However, Dodge [58], Hines and Maddox [59] have provided techniques for reducing the number of trials that is required.

19.14.1 Dew Point Calculations

The dew point of a vapor is that combination of temperature and pressure at which the first drop of the liquid condenses. The dew point criterion can be expressed as:

$$\sum x_i = \sum \frac{y_i}{K_i} = 1.0 \tag{19.110}$$

For binary mixtures of components 1 and 2, then

$$y_i = \frac{K_1(1-K_2)}{K_1-K_2} \tag{19.111}$$

and

$$y_2 = 1.0 - y_1 \tag{19.112}$$

These two equations can be used to calculate the composition of a binary vapor that will begin to condense at a given temperature and pressure.

The Algorithm

Bubble Point Calculation for the first trace of vapor of a component mixture x_i is known

If P_{Tot} is given, guess T to satisfy $P_{Tot} = \sum x_i\, p_i^o(T)$

Then, $y_i = \dfrac{x_i\, p_i^o}{P_{Tot}}$

where

x_i = mole fraction of component i in the liquid phase.
y_i = mole fraction of component i in the vapor phase
p_i^o = vapor pressure of the component i at the working temperature
P_{Tot} = π = total pressure of the mixture, in psia (psia = psig + 14.7), in bara (bara = barg + 1.013)
T = working temperature, °F (°C).

Dew Point Calculation

The last trace of liquid, when the entire solution is boiled off.

y_i is known
Use relation: $\sum x_i = 1.0$

If P_{Tot} is known, guess T to satisfy

$$\sum \frac{y_i\, P_{Tot}}{p_i^o(T)} = 1.0$$

If T is known, get $P_{Tot} = \dfrac{1}{\sum \dfrac{y_i}{p_i^o(T)}}$

Then $y_i = \dfrac{x_i\, p_i^o}{P_{Tot}}$

Example 19.6: Dew-Point Temperature Calculations for a Multicomponent Mixture

A multicomponent hydrocarbon gas mixture is composed as follows:

Component	Mol %	K
Methane (CH_4)	27.52	7.88
Ethane (C_2H_6)	16.34	2.77
Propane (C_3H_8)	29.18	1.18
Isobutane (iC_4H_{10})	5.37	0.61
Normal butane (nC_4H_{10})	17.18	0.48
Isopentane (iC_5H_{12})	1.72	0.264
Normal pentane (nC_5H_{12})	2.18	0.225
Hexane (C_6H_{14})	0.47	0.102
Heptane (C_7H_{16})	0.04	0.048

Operating temperature and pressure are 178°C and 400 psia. Determine the dew point temperature of the gas mixture.

Solution

The Excel spreadsheet program (Example 19.6.xlsx) calculates the dew and bubble points of any multicomponent hydrocarbon mixture based on the user supplied K-values (see Figures 19.12a–19.12d). The program will handle feed streams containing up to 15 components. The feed entries may be moles, mole fraction or mole percent. The dew point $\left(\sum x = \sum y/K\right)$ should be 1.0 or very close to 1.0, depending on the acceptable tolerance. If the sum obtained is not equal to 1.0 (within the user tolerance), a message is shown instructing the user to assume a new temperature. The iteration is repeated using the K_i's corresponding to the new temperature selected. The temperature supplied does not affect the computation, but is printed and stored as a reminder to the user of which temperature (and corresponding K-values) was used for the last iteration. Table 19.13 shows the snapshot and results of the Excel spreadsheet calculations and results of Example 19.6.

Table 19.13 Excel spreadsheet results for dew and bubble points calculations.

Dew point calculations		**At 178 °F**		**400 psia**		
Number of comp.	**Feed comp. mole %**	**K-value**	**Number of comp.**	**Mole fraction, y**	**x = y/K**	
1	27.52	7.88	1	0.2752	0.0349	
2	16.34	2.77	2	0.1634	0.059	
3	29.18	1.18	3	0.2918	0.2473	
4	5.37	0.61	4	0.0537	0.088	
5	17.18	0.48	5	0.1718	0.3579	
6	1.72	0.264	6	0.0172	0.0652	
7	2.18	0.225	7	0.0218	0.0969	
8	0.47	0.102	8	0.0047	0.0461	
9	0.04	0.048	9	0.0004	0.0083	
Total	100			1	1.0036	Total Dew Point
				Raise the Dew Point Temperature		

Raise the dew point temperature of 178°F

Example 19.7: Bubble Point Temperature Calculations for Multicomponent Mixture

At 200 psia, calculate the bubble point temperature of the following:

Component	Mol %	K @ 260°F	K @ 235°F	K @ 237°F
iC_4H_{10}	18.2	1.92	1.62	1.65
nC_4H_{10}	23.8	1.58	1.35	1.35
iC_5H_{12}	33.7	0.93	0.76	0.77
nC_5H_{12}	12.1	0.81	0.64	0.64
C_6H_{14}	12.2	0.42	0.315	0.32

Solution

The Excel spreadsheet program (Example 19.7.xlsx) determines the bubble points of the hydrocarbon mixture. As with the dew point, the bubble point $\left(\sum y = \sum K x\right)$ is computed until the sum is 1.0 or close to this value, obtained depending on the accuracy required. If the sum printed is not equal to 1.0 (within tolerance), the iteration is repeated using the K_i's corresponding to the new temperature. Tables 19.14, 19.15, and 19.16 show the snapshots of Excel spreadsheet calculations for the bubble points of the hydrocarbon mixture at 260°F, 235°F, and 237°F, respectively. By interpolation, the bubble temperature is 237.3°F at 200 psia.

Table 19.14 Excel spreadsheet results for dew and bubble points calculations.

Bubble point calculations		**At 260 °F**		**200 psia**		
Number of comp.	**Feed comp. mole %**	**K-value**	**Number of comp.**	**Mole fraction, x**	**y = Kx**	
1	18.2	1.92	1	0.182	0.3494	
2	23.8	1.58	2	0.238	0.376	
3	33.7	0.93	3	0.337	0.3134	
4	12.1	0.81	4	0.121	0.098	
5	12.2	0.42	5	0.122	0.0512	
Total	100			1	1.188	Total Bubble Point
					Lower the Bubble Point	

Lower the bubble point temperature of 260°F

Table 19.15 Excel spreadsheet results for dew and bubble points calculations.

Bubble point calculations		At 235 °F		200 psia		
Number of comp.	**Feed comp. mole %**	**K-value**	**Number of comp.**	**Mole fraction, x**	**y = Kx**	
1	18.2	1.62	1	0.182	0.2948	
2	23.8	1.35	2	0.238	0.3213	
3	33.7	0.76	3	0.337	0.2561	
4	12.1	0.64	4	0.121	0.0774	
5	12.2	0.315	5	0.122	0.0384	
Total	100			1	0.988	Total Bubble Point
					Raise the Bubble Point Temperature	

Raise the bubble point temperature of 235°F

Table 19.16 Excel spreadsheet results for dew and bubble points calculations.

Bubble Point Calculations		At 237 °F		200 psia		
Number of comp.	**Feed comp. mole %**	**K-value**	**Number of comp.**	**Mole fraction, x**	**y = Kx**	
1	18.2	1.65	1	0.182	0.3003	
2	23.8	1.35	2	0.238	0.3213	
3	33.7	0.77	3	0.337	0.2595	
4	12.1	0.64	4	0.121	0.0774	
5	12.2	0.32	5	0.122	0.039	
Total	100			1	0.9975	Total Bubble Point
					Raise the Bubble Point Temperature	

Raise the bubble point temperature of 237°F

19.15 Equilibrium Flash Computations

Flash vaporization calculations involving multicomponent mixtures are necessary for numerous processes, often to determine the condition of the feed to a fractionating column or to determine the flow of vapor from reboiler or condenser.

Such calculations often involve trial-and-error solutions, which are time consuming, tedious and subject to error if performed manually. They are best carried out with computer-aided, process design and simulation programs, such as UniSim® (Honeywell) Design, Aspen Plus, HYSYS, ChemCad, and Pro II. This author [60] has developed an improved iterative convergence method first suggested by Oliver [61] and later modified by Kostecke [62], for isothermal equilibrium flash computations. Multicomponent flash (isothermal and adiabatic) computations are now incorporated as part of overall process simulation and equipment design package. However, single-stage flash fractionation processes are also employed to separate the light components in a feed and as a preliminary step before a multicomponent fractionation column, e.g., crude oil distillation. Table 19.17 lists ways of producing two-phase mixtures from a single phase at appropriate conditions. The last process in the table typifies the well head separation that takes place in an oil field. It is termed a flash process because the vapor forms due to the rapid drop in the pressure.

19.15.1 Fundamentals

To carry out an appropriate flash calculation, the pressure, P and the temperature, T, must be known. If the values of P and T in the separating vessel are fixed, the value of P must not be so high that the two phases cannot exist at any value of T. Nor must T lie outside the bubble point and dew point range corresponding to P. For a valid two-phase equilibrium calculation, the following relationship must be satisfied

$$T_{bp} < T_s < T_{dp} \tag{19.113}$$

where

T_{bp} = the bubble point temperature
T_s = the specified temperature
T_{dp} = the dew point temperature

The existence of a valid two-phase flash can be verified with the design equations for the bubble point, dew point, and equilibrium data calculated at the specified pressure and temperature.

Table 19.17 Ways to produce two-phase mixture.

Initial	Action to produce two-phase mixture
Gas	Cool, possibly after initial compression
Gas	Expand through a valve or an engine
Liquid	Heat to achieve partial vaporization
Liquid	Reduce pressure through a valve, if close to saturation

19.15.2 Calculation of Bubble Point and Dew Point

The design equations for the bubble point and dew point are:

Bubble point

$$f_1 = \sum_{i=1}^{n} K_i\, x_i \tag{19.114}$$

Dew point

$$f_2 = \sum_{i=1}^{n} y_i / K_i \tag{19.115}$$

The calculation method is as follows (at constant pressure):

1. Guess a temperature, T.
2. Calculate K values.
3. Calculate the sum of the right side of Equation 19.114 for bubble point calculation. If < 1.0, increase the temperature. If >1.0, decrease the temperature. Repeat steps 2 and 3 until convergence is attained.
4. For dew point calculation, obtain the right side of Equation 19.115. If <1.0, decrease the temperature. If >1.0, increase the temperature. Again, repeat steps 2 and 4 until a convergence is reached.

Table 19.18 illustrates the phase condition using the light-vapor data associated with the specified pressure and temperature. From Table 19.18, the conditions for a valid two-phase equilibrium flash are:

$$f_1 = \sum_{i=1}^{n} K_i\, n_i > 1.0 \tag{19.116}$$

and

$$f_2 = \sum_{i=1}^{n} n_i / K_i > 1.0 \tag{19.117}$$

Table 19.18 Equilibrium flash criteria.

	$f_1 = \sum_{i=1}^{n} K_i n_i$	$f_2 = \sum_{i=1}^{n} n_i / K_i$
Subcooled liquid	<1	>1
Bubble point	=1	>1
Two-phase condition	>1	>1
Dew point	>1	=1
Superheated vapor	>1	<1

Feed analyses of component concentrations are generally unavailable for complex hydrocarbon mixtures with a final boiling point of > 38°C, unless the feed is sub-divided into pseudo-components (narrow boiling fractions). This sub-division enables the mole fraction and equilibrium constant, K, to be estimated, and consequently, flash calculation of the mixture can be carried out for the mixture. A source of K values for light hydrocarbons systems is the DePriester [63] charts, which are shown in Figures 19.12a to 19.12d. These give the K values over a wide range of temperature and pressure. Because the charts are nomographs, a straight edge connecting the temperature and pressure for which K values are required will intersect curves for each compound at its K values. The values have been obtained by calculating fugacities from an equation of state. Use of these charts for exact determination of K values requires trial-and-error calculations. Hadden and Grayson [64] have presented correlations for hydrocarbon vapor-liquid distribution ratios, and the charts from their work are shown in Figures 19.27 and 19.28. These charts can be readily used for determining the vapor–liquid distribution coefficients. Figure 19.29 shows a continuous equilibrium flash fractionation process.

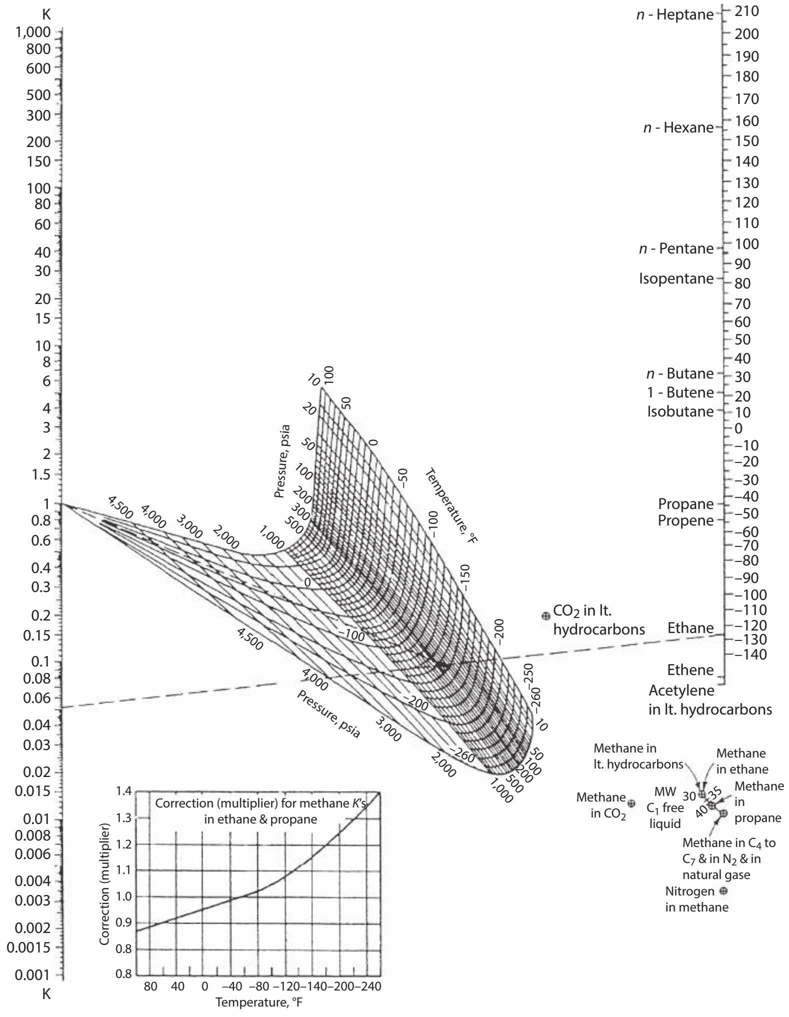


Figure 19.27 Vapor–liquid equilibrium constants, −260 to +100°F (source: Hadden and Grayson [270]).

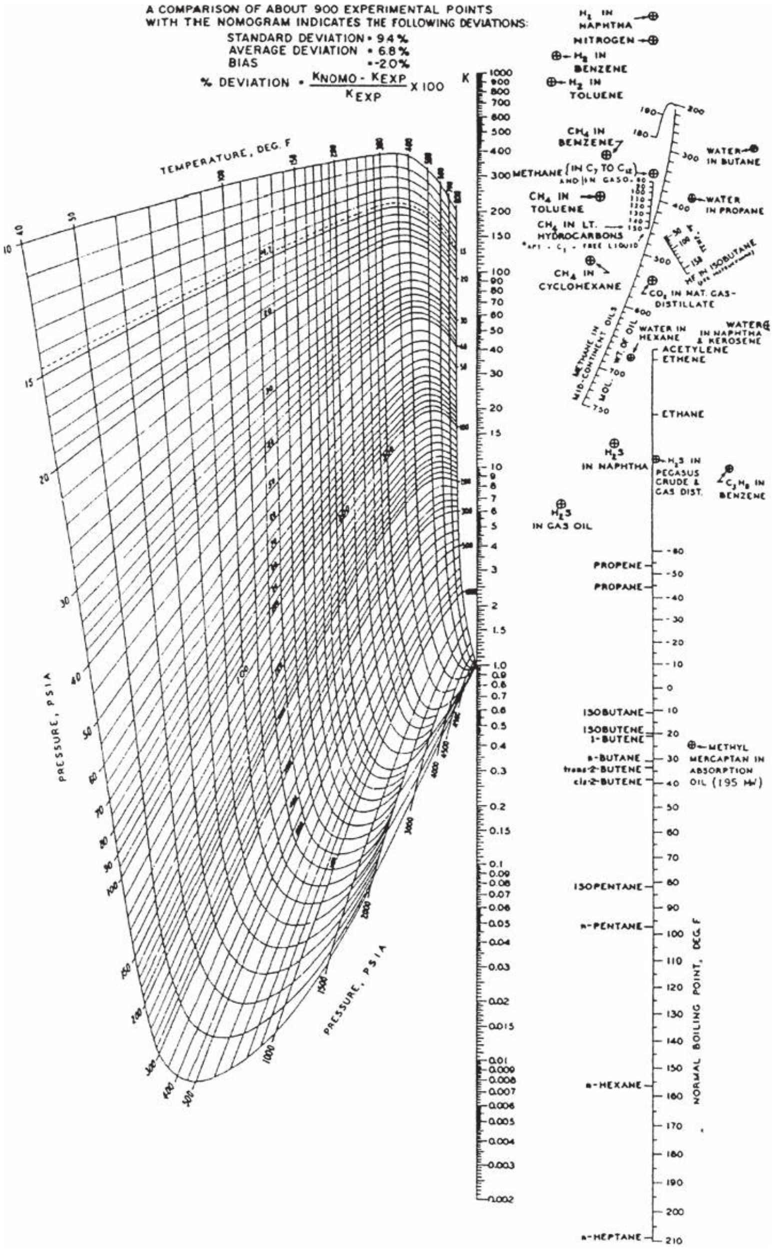


Figure 19.28 Vapor–liquid equilibrium constants, 40 to 800°F (source: Hadden and Grayson [270]).

The Equations

The following equations are used for the multicomponent equilibrium flash calculations.

$$f_1 = \sum_{i=1}^{n} n_i K_i < 1 \text{ all liquid} \tag{19.118}$$

$$f_2 = \sum_{i=1}^{n} n_i / K_i < 1 \text{ all vapor} \tag{19.119}$$

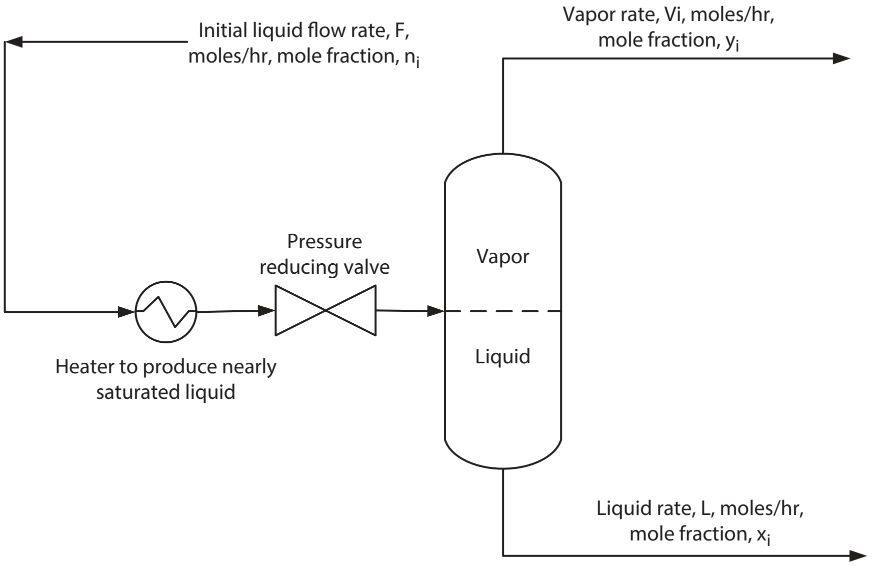


Figure 19.29 Continuous equilibrium flash fractionation.

$$C_i = \frac{[M_i(K_i - 1)(R + 1)]}{(K_i + R)} \tag{19.120}$$

where

R = the liquid–vapor (L/V) ratio. The new L/V ratio for each iterative calculation, R′, is determined from:

$$R' = \frac{[FR - E(R + 1)]}{[F + E(R + 1)]} \tag{19.121}$$

The constants E and F are determined by

$$E = \sum_{i=1}^{n} C_i \tag{19.122}$$

and

$$F = \sum_{i=1}^{n} (C_i)^2 / M_i \tag{19.123}$$

The Algorithm

The computation then determines where |E| < 0.001. If |E| is not less than 0.001, R is set equal to R′ and the calculations of Equations 19.120, 19.121, 19.122, and 19.123 are repeated until |E| < 0.001.

Once |E| < 0.001, the calculation proceeds to find the total moles of component i in both the liquid and vapor phases, that is L_i and V_i. These are given by

$$L_i = \frac{M_i R}{(K_i + R)} \tag{19.124}$$

and

$$V_i = M_i - L_i \tag{19.125}$$

The mole fractions of components in the feed, the liquid and vapor phases are evaluated as follows:

$$n_i = \frac{M_i}{\sum_{i=1}^{n} M_i} \tag{19.126}$$

$$x_i = \frac{L_i}{\sum_{i=1}^{n} L_i} \tag{19.127}$$

and

$$y_i = \frac{V_i}{\sum_{i=1}^{n} V_i} \tag{19.128}$$

where

$$\sum_{i=1}^{n} n_i = \sum_{i=1}^{n} x_i = \sum_{i=1}^{n} y_i = 1 \tag{19.129}$$

where

E, F = constants
K_i = equilibrium flash constant for each component in the feed stream
L_i = total moles of component i in the liquid phase
L = total moles of liquid at equilibrium conditions
M_i = total moles of component i in the feed stream
n_i = mole fraction of component i in the feed stream
R = liquid–vapor (L/V) ratio
R′ = new L/V ratio for each iteration calculation
V = total moles of vapor at equilibrium conditions
V_i = total moles of component i in the vapor phase
x_i = mole fraction of component i in the liquid phase
y_i = mole fraction of component i in the vapor phase

The method described here is based on the vapor–liquid equilibrium relationships given in handbooks available from the Gas Processors Suppliers Association [28].

The Program

An Excel spreadsheet program (Example 19.8.xlsx) has been developed for the flash calculation; it is based upon the vapor–liquid equilibrium relationships given in Engineering Data books available from the Gas Processors Suppliers Association, Tulsa and other literature. It will handle calculations with feed streams containing up to 15 components. In addition, the calculation will check the feed composition at flash conditions for dew and bubble points (i.e., whether the feed is either all vapor or all liquid). These checks are performed before the flash calculations are started. If the feed is above the dew point or below the bubble point, an appropriate message is displayed on the screen. A default value of R (L/V ratio) = 1 is incorporated in the program to start the iterative process. The program is not applicable to an adiabatic flash process when the feed stream is at pressure higher than the flash pressure, and the heat of vaporization is provided by the enthalpy of the feed. In this situation, the flash temperature will be unknown and must be obtained by trial and error. A temperature must be found at which both the material and energy balances are satisfied.

Example 19.8: Determine the Amount of liquid and Vapor in a Multicomponent Mixture

Determine the amount of liquid and vapor in a nine-component still product accumulator stream that is compressed to 370 psia and cooled to 90°F.

Process Data

Number	Component	Feed moles/h	Equilibrium constant, K
1	CH_4	2752	7.2
2	C_2H_6	1634	1.65
3	C_3H_8	2918	0.54
4	iC_4H_{10}	537	0.25
5	nC_4H_{10}	1718	0.185
6	iC_5H_{12}	172	0.088
7	nC_5H_{12}	218	0.069
8	C_6H_{14}	47	0.028
9	C_7H_{16}	4	0.00078

Solution

The Excel spreadsheet program Example 19.8.xlsx calculates the liquid and vapor streams from the equilibrium constants and the molar flow rates of the components at the supplied temperature and pressure. Table 19.19 shows the snapshots and results of Example 19.8.

Example 19.9: Determination of the Flow Rates of Liquid and Vapor in a Gas Liquefaction Plant

A feed stream of natural gas is flashed at 600 psia and 20°F in a gas liquefaction plant. Determine the flow rates of the liquid and vapor streams of a feed flow rate of 1000 moles/h.

Table 19.19 Multicomponent equilibrium flash calculations of Example 19.8.

Multicomponent equilibrium											
Flash Calculation At 90 °F 370 psia											
			Liquid to Vapor Ratio (L/V)				**1.18298 R**				
Component Number	**K-value**	**Feed mole/h**	**Mole faction, x**	**f_1**	**f_2**	**Mi x*F**	**Ci**	**Li moles/h**	**x, Mole. Fiac**	**Yi moles/h**	**y, Mole. Fiac**
CH_4	7.2	2752	0.2752	1.98144	0.0382	2752.00	4443.15	388.35	0.07	2363.65	0.516
C_2H_6	1.65	1634	0.1634	0.26961	0.099	1634.00	818.41	682.32	0.13	951.68	0.2078
C_3H_8	0.54	2918	0.2918	0.157572	0.5404	2918.00	-1700.64	2003.47	0.37	914.53	0.1996
iC_4H_{10}	0.25	537	0.0537	0.013425	0.2148	537.00	-613.54	443.31	0.08	93.69	0.0205
nC_4H_{10}	0.185	1718	0.1718	0.031783	0.9286	1718.00	-2234.35	1485.66	0.27	232.34	0.0507
iC_5H_{12}	0.088	172	0.0172	0.0015136	0.1955	172.00	-269.42	160.09	0.03	11.91	0.0026
nC_5H_{12}	0.069	218	0.0218	0.0015042	0.3159	218.00	-353.88	205.99	0.04	12.01	0.0026
nC_6H_{14}	0.028	47	0.0047	0.0001316	0.1679	47.00	-82.35	45.91	0.01	1.09	0.0002
nC_7H_{16}+heavier	0.0008	4	0.0004	3.12E-07	0.5128	4.00	-7.37	4.00	0.00	0.00	6E-07
Total		10000	1	2.46	3.01		0.00	5419.10	1.00	4580.90	1
				>1	>1						
1				two-**phase**							

$$f_1 = \sum_{i=1}^{n} n_i K_i < 1 \text{ all liquid} \tag{9.118}$$

$$f_2 = \sum_{i=1}^{n} n_i / K_i < 1 \text{ all vapor} \tag{19.119}$$

$$C_i = \frac{[M_i(K_i - 1)(R+1)]}{(K_i + R)} \tag{19.120}$$

$$R' = \frac{[FR - E(R+1)]}{[F + E(R+1)]} \tag{19.121}$$

$$E = \sum_{i=1}^{n} C_i \tag{19.122}$$

$$F = \sum_{i=1}^{n} (C_i)^2 / M_i \tag{19.123}$$

$$L_i = \frac{M_i R}{(K_i + R)} \tag{19.124}$$

Process Data

Number	Component	Feed Moles fraction	Equilibrium constant, K
1	CO_2	0.0112	0.90
2	CO_4	0.8957	2.70
3	C_2H_6	0.0526	0.38
4	C_3H_8	0.0197	0.098
5	iC_4H_{10}	0.0068	0.038
6	nC_4H_{10}	0.0047	0.024
7	C_5H_{12}	0.0038	0.0075
8	nC_6H_{14}	0.0031	0.0019
9	nC_7H_{16} and heavier	0.0024	0.0007

Solution

Table 19.20 shows the Excel spreadsheet snapshots and results of isothermal equilibrium flash calculation at 600 psia.

Table 19.20 Multicomponent equilibrium flash calculations of Example 19.9.

Multicomponent equilibrium Flash Calculation At 20 °F 600 psia										
				Liquid to Vapor Ratio (L/V)			**0.042569479**	**R**		
Number of Comp.	**K-value**	**Mole Fraction, x**	**f_1**	**f_2**	**Mi x*F**	**Ci**	**Li moles/h**	**x_i Mole. Frac**	**Vi moles/h**	**y_i Mole. Frac**
CO_2	0.9	0.0112	0.01008	0.012444	11.20	-1.24	0.51	0.01	10.69	0.0111494
CH_4	2.7	0.8957	2.41839	0.331741	895.70	578.84	13.90	0.34	881.80	0.9193348
C_2H_6	0.38	0.0526	0.019988	0.138421	52.60	-80.46	5.30	0.13	47.30	0.0493147
C_3H_8	0.098	0.0197	0.0019306	0.20102	19.70	-131.79	5.97	0.15	13.73	0.0143188
iC_4H_{10}	0.038	0.0068	0.0002584	0.178947	6.80	-84.65	3.59	0.09	3.21	0.0033437
nC_4H_{10}	0.024	0.0047	0.0001128	0.195833	4.70	-71.84	3.01	0.07	1.69	0.0017666
C_5H_{12}	0.0075	0.0038	0.0000285	0.506667	3.80	-78.53	3.23	0.08	0.57	0.0005934
nC_6H_{14}	0.0019	0.0031	0.00000589	1.631579	3.10	-72.54	2.97	0.07	0.13	0.0001381
nC_7H_{16}+heavier	0.0007	0.0024	0.00000168	3.428571	2.40	-57.79	2.36	0.06	0.04	4.048E-05
Total		1	2.45	6.63		0.00	40.83	1.00	959.17	1
			>1	>1						
			two-phase							

19.16 Degrees of Freedom

The design of separating systems involves many parameters, and the number of variables that can be arbitrarily specified is known as the degrees of freedom. This is determined by subtracting the number of thermodynamic equilibrium equations from the number of variables. For non-reacting systems, the Gibbs phase rule is defined by:

$$F = C - P + 2 \tag{19.130}$$

where

F = degrees of freedom (i.e., the number of independent properties that have to be specified to determine all the intensive properties of each phase of the system of interest).

C = number of components in the system. For circumstances that involve chemical reactions, C is not identical to the number of chemical compounds in the system but is equal to the number of chemical compounds less the number of independent-reaction and other equilibrium relationships among these compounds.

P = number of phases that can exist in the system. A phase is a homogeneous quantity of material, e.g., gas, pure liquid, a solution, or a homogeneous solid.

Variables of the kind with which the phase rule is concerned are termed phase-rule variables, and they are intensive properties of the system, i.e., properties that do not depend on the quantity of material present.

For example, in a binary system such as methanol–water having liquid and vapor phases,

$$C = 2,\ P = 2 \text{ and}$$

$$F = 2 - 2 + 2 = 2$$

When temperature and pressure are set, all the degrees of freedom are used, and at equilibrium all compositions are determined from the experiment. Alternatively, the pressure and components' mole fractions (methanol x_{Mtoh}, and water, x_w) are known, the temperature and the other mole fractions are determined. The phase rule refers to intensive variables where temperature, pressure, specific (per unit mass) values (i.e., specific volume) or mole fractions do not depend on the total amount of material present. The extensive variables, i.e., number of moles, flow rate and volume depend on the amount of material and are not included in the degree of freedom.

Another example is the ideal gas equation PV = nRT, where if the phase rule is applied for a single phase and for a pure gas, P = 1, C = 1, then F = 1 − 1 + 2 = 2 variables to be specified. Since the phase rule is concerned with intensive properties only, the following phase-rule variables in the ideal gas law:

$$\left.\begin{array}{l} P \\ \hat{V} \text{ (specific molar volume)} \\ T \end{array}\right\} \text{3 intensive properties}$$

By specifying two intensive variables (F = 2), the third variable can be determined.

When a set of independent equations representing the material balances for a problem is prepared, the number of variables whose values are unknown is counted. If more variables whose values are unknown exist than independent equations, then an infinite number of solutions exist for a material balance, and such problems are referred to as *underspecified (underdetermined)*. Either values of additional variables must be found to make up the deficit or the problem must be posed as an optimization problem. Alternatively, if fewer values of the variables whose values are unknown exist than the independent equations, the problem is *overspecified (overdetermined)* and no solution exists to the problem, showing that the equations are inconsistent. In this instance, the problem might be posed as an optimization problem to minimize the sum of the squares of the deviations of the equations from zero (or their right-hand constants).

The difference between the number of variables whose values are unknown and the number of independent equations is termed the number of degrees of freedom. If the degrees of freedom are positive, such as 2, two additional independent equations or specifications of variables must be sought to provide a unique solution to the material balance problem. Conversely, if the degrees of freedom are negative e.g. -1, there are many equations or not enough variables in the problem. *Zero degrees of freedom means that the material balances problem is properly specified*, and the equations can be solved for the variables whose values are unknown. In this instance, the number of degrees of freedom is the number of variables in a set of independent equations to which values must be assigned so that the equations can be solved. This can be expressed by

$$N_d = N_v - N_r \tag{19.131}$$

where

N_d = number of degrees of freedom
N_v = number of variables
N_r = number of equations

19.17 UniSim (Honeywell) Software

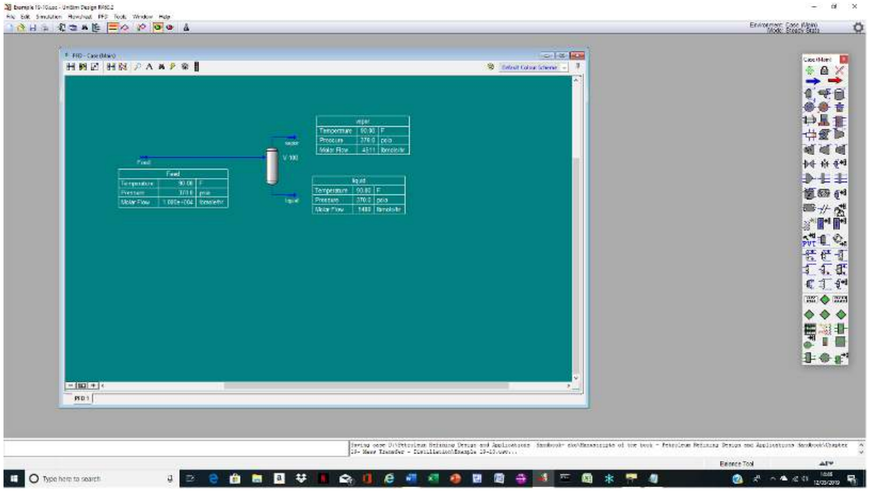
UniSim® Design R460.2 simulation software is an intuitive and interactive process modeling package that allows engineers to create steady-state and dynamic models for plant design, performance monitoring, troubleshooting, operational improvement, business planning and asset management. Further, it provides users with Honeywell's domain expertise in process simulation and operator training. Using industry-specific unit operating models and powerful tools to optimize operating parameters for feedstock changes, UniSim Design allows the steady-state simulation of oil and gas, refining, chemical and petrochemical processes.

In this book, UniSim Design simulation software is employed to verify some of the design calculations.

Example 19.10: Application of UniSim Design Software for Example 19.8

Example 19.8 is repeated using UniSim Design R460.2 software (Example 19.10.usc) to determine the amount of liquid and vapor in a nine-component still product accumulator stream that is compressed to 370 psia and cooled to 90°F. Figure 19.30 shows the screen shot of UniSim Design 460.2.

Table 19.21 shows the simulation results from Honeywell UniSim when the Peng-Robinson equation for vapor–liquid equilibrium (VLE) is used. The results for the total vapor and liquid flow rates give good agreement with the results in Table 19.19.



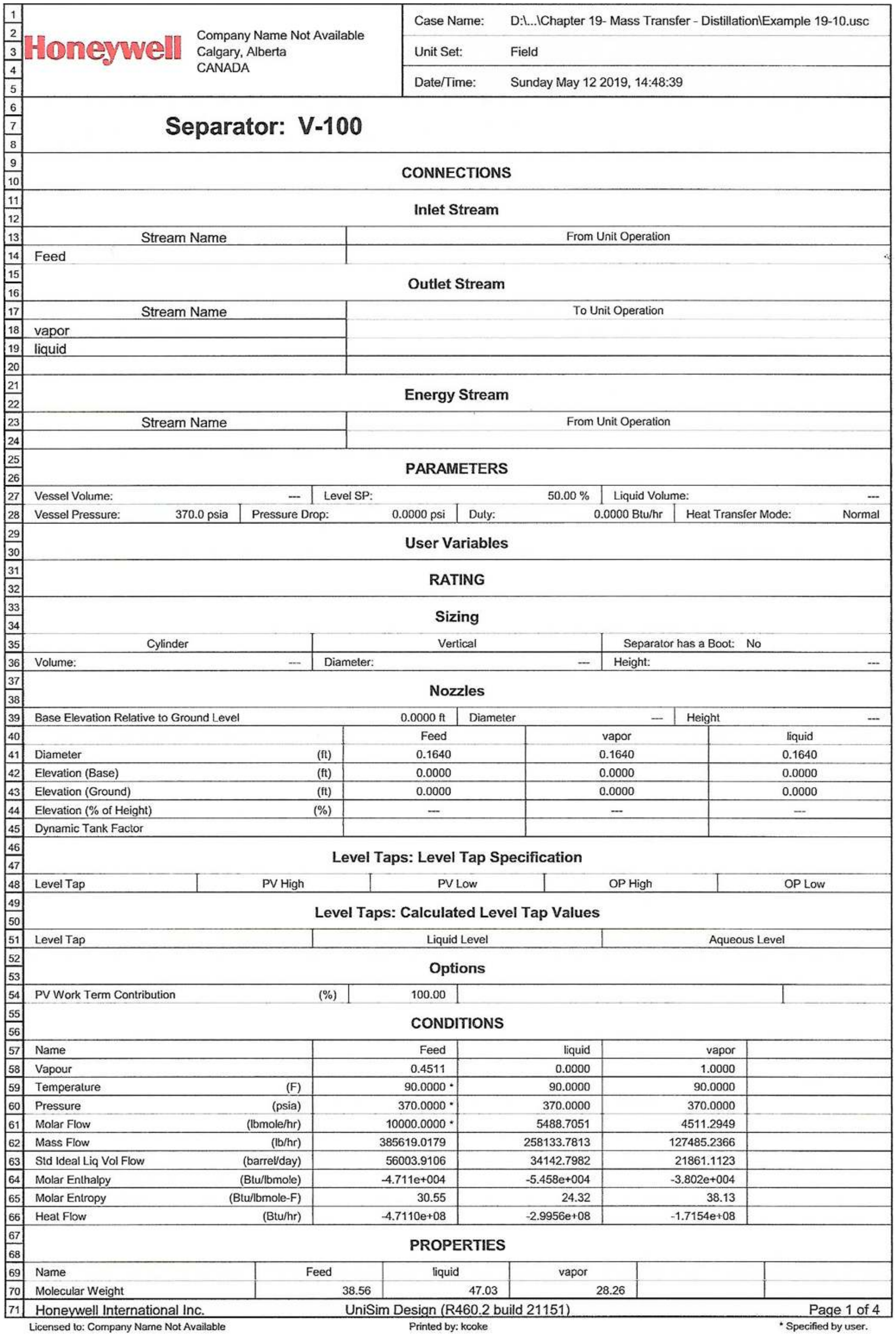
Feed		
Temperature	90.00	F
Pressure	370.0	psia
Molar Flow	1.000e+004	lbmole/hr

vapor		
Temperature	90.00	F
Pressure	370.0	psia
Molar Flow	4511	lbmole/hr

liquid		
Temperature	90.00	F
Pressure	370.0	psia
Molar Flow	1488	lbmole/hr

Figure 19.30 A screen-shot of the process flow diagram of a flash drum of Example 19.10 (courtesy of Honeywell Process Solution. UniSim® Design R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International, Inc.).

Table 19.21 Results of the simulation using UniSim® Design R460.2 Software for Example 19.10.



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	D:\...\Chapter 19- Mass Transfer - Distillation\Example 19-10.usc
Unit Set:	Field
Date/Time:	Sunday May 12 2019, 14:48:39

Separator: V-100

CONNECTIONS

Inlet Stream

Stream Name	From Unit Operation
Feed	

Outlet Stream

Stream Name	To Unit Operation
vapor	
liquid	

Energy Stream

Stream Name	From Unit Operation

PARAMETERS

Vessel Volume:	---	Level SP:		50.00 %	Liquid Volume:		---
Vessel Pressure:	370.0 psia	Pressure Drop:	0.0000 psi	Duty:	0.0000 Btu/hr	Heat Transfer Mode:	Normal

User Variables

RATING

Sizing

Cylinder		Vertical		Separator has a Boot:	No
Volume:	---	Diameter:	---	Height:	---

Nozzles

Base Elevation Relative to Ground Level	0.0000 ft	Diameter	---	Height	---

		Feed	vapor	liquid
Diameter	(ft)	0.1640	0.1640	0.1640
Elevation (Base)	(ft)	0.0000	0.0000	0.0000
Elevation (Ground)	(ft)	0.0000	0.0000	0.0000
Elevation (% of Height)	(%)	---	---	---
Dynamic Tank Factor				

Level Taps: Level Tap Specification

Level Tap	PV High	PV Low	OP High	OP Low

Level Taps: Calculated Level Tap Values

Level Tap	Liquid Level	Aqueous Level

Options

PV Work Term Contribution	(%)	100.00	

CONDITIONS

Name		Feed	liquid	vapor	
Vapour		0.4511	0.0000	1.0000	
Temperature	(F)	90.0000 *	90.0000	90.0000	
Pressure	(psia)	370.0000 *	370.0000	370.0000	
Molar Flow	(lbmole/hr)	10000.0000 *	5488.7051	4511.2949	
Mass Flow	(lb/hr)	385619.0179	258133.7813	127485.2366	
Std Ideal Liq Vol Flow	(barrel/day)	56003.9106	34142.7982	21861.1123	
Molar Enthalpy	(Btu/lbmole)	-4.711e+004	-5.458e+004	-3.802e+004	
Molar Entropy	(Btu/lbmole-F)	30.55	24.32	38.13	
Heat Flow	(Btu/hr)	-4.7110e+08	-2.9956e+08	-1.7154e+08	

PROPERTIES

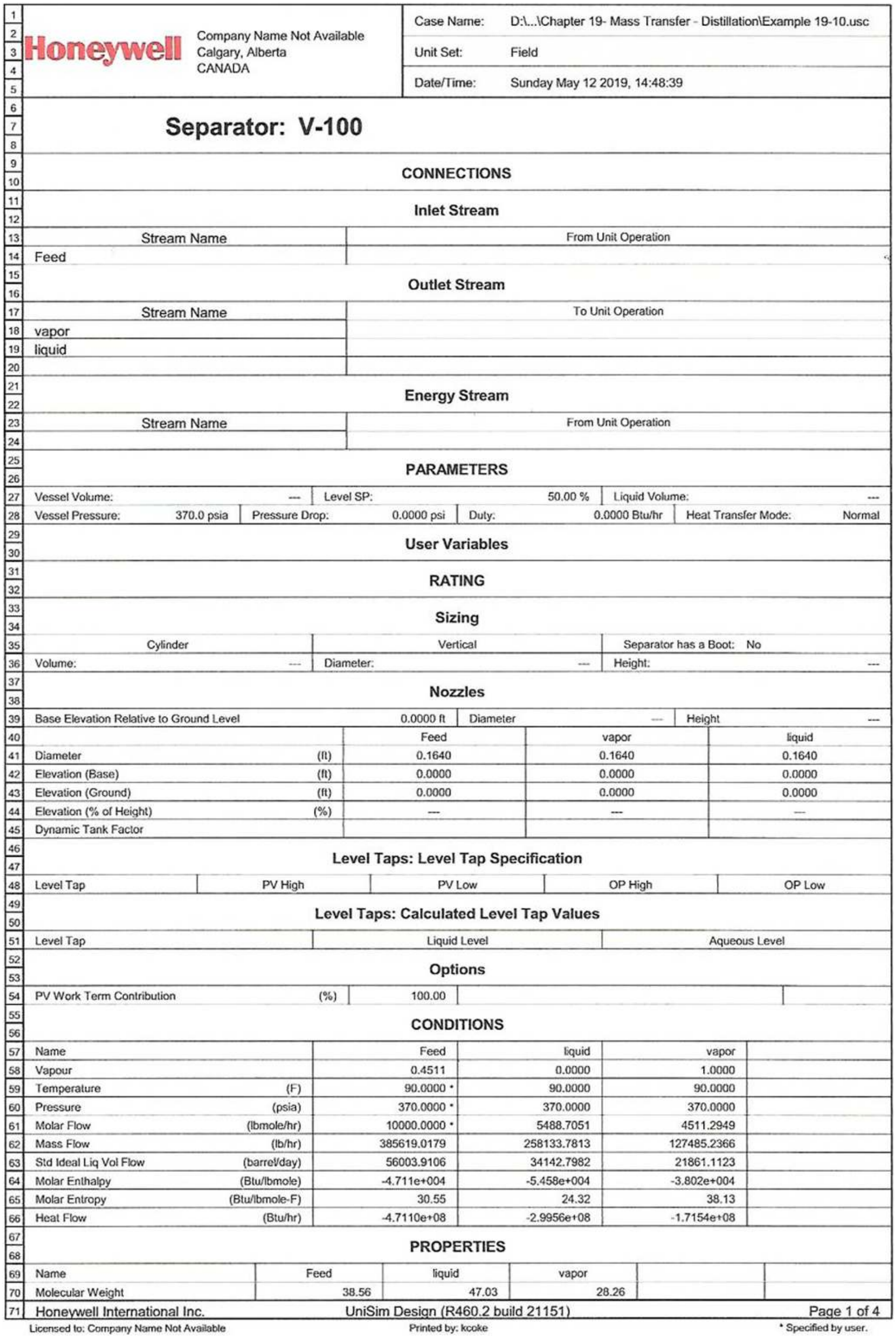
Name	Feed	liquid	vapor		
Molecular Weight	38.56	47.03	28.26		

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 1 of 4

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

(*Continued*)

Table 19.21 Results of the simulation using UniSim® Design R460.2 Software for Example 19.10. (*Continued*)



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	D:\...\Chapter 19- Mass Transfer - Distillation\Example 19-10.usc
Unit Set:	Field
Date/Time:	Sunday May 12 2019, 14:48:39

Separator: V-100

CONNECTIONS

Inlet Stream

Stream Name	From Unit Operation
Feed	

Outlet Stream

Stream Name	To Unit Operation
vapor	
liquid	

Energy Stream

Stream Name	From Unit Operation

PARAMETERS

Vessel Volume:	---	Level SP:	50.00 %	Liquid Volume:	---		
Vessel Pressure:	370.0 psia	Pressure Drop:	0.0000 psi	Duty:	0.0000 Btu/hr	Heat Transfer Mode:	Normal

User Variables

RATING

Sizing

Cylinder		Vertical		Separator has a Boot:	No
Volume:	---	Diameter:	---	Height:	---

Nozzles

Base Elevation Relative to Ground Level	0.0000 ft	Diameter	---	Height	---

		Feed	vapor	liquid
Diameter	(ft)	0.1640	0.1640	0.1640
Elevation (Base)	(ft)	0.0000	0.0000	0.0000
Elevation (Ground)	(ft)	0.0000	0.0000	0.0000
Elevation (% of Height)	(%)	---	---	---
Dynamic Tank Factor				

Level Taps: Level Tap Specification

Level Tap	PV High	PV Low	OP High	OP Low

Level Taps: Calculated Level Tap Values

Level Tap	Liquid Level	Aqueous Level

Options

PV Work Term Contribution	(%)	100.00

CONDITIONS

Name		Feed	liquid	vapor
Vapour		0.4511	0.0000	1.0000
Temperature	(F)	90.0000 *	90.0000	90.0000
Pressure	(psia)	370.0000 *	370.0000	370.0000
Molar Flow	(lbmole/hr)	10000.0000 *	5488.7051	4511.2949
Mass Flow	(lb/hr)	385619.0179	258133.7813	127485.2366
Std Ideal Liq Vol Flow	(barrel/day)	56003.9106	34142.7982	21861.1123
Molar Enthalpy	(Btu/lbmole)	-4.711e+004	-5.458e+004	-3.802e+004
Molar Entropy	(Btu/lbmole-F)	30.55	24.32	38.13
Heat Flow	(Btu/hr)	-4.7110e+08	-2.9956e+08	-1.7154e+08

PROPERTIES

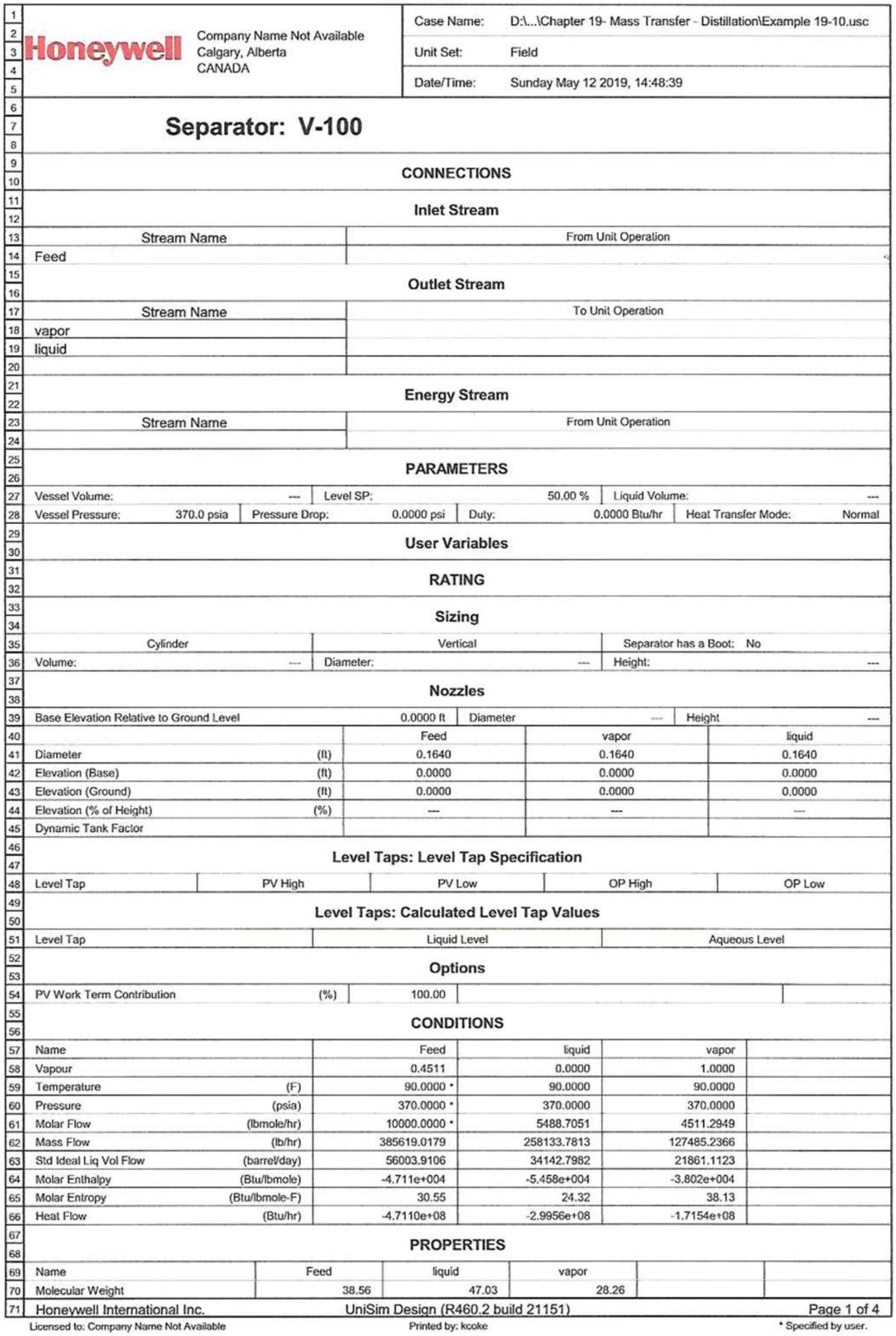
Name	Feed	liquid	vapor
Molecular Weight	38.56	47.03	28.26

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 1 of 4

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

(*Continued*)

Table 19.21 Results of the simulation using UniSim® Design R460.2 Software for Example 19.10. (*Continued*)



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	D:\...\Chapter 19- Mass Transfer - Distillation\Example 19-10.usc
Unit Set:	Field
Date/Time:	Sunday May 12 2019, 14:48:39

Separator: V-100

CONNECTIONS

Inlet Stream

Stream Name	From Unit Operation
Feed	

Outlet Stream

Stream Name	To Unit Operation
vapor	
liquid	

Energy Stream

Stream Name	From Unit Operation

PARAMETERS

Vessel Volume:	---	Level SP:	50.00 %	Liquid Volume:	---		
Vessel Pressure:	370.0 psia	Pressure Drop:	0.0000 psi	Duty:	0.0000 Btu/hr	Heat Transfer Mode:	Normal

User Variables

RATING

Sizing

Cylinder		Vertical		Separator has a Boot:	No
Volume:	---	Diameter:	---	Height:	---

Nozzles

Base Elevation Relative to Ground Level	0.0000 ft	Diameter	---	Height	---

		Feed	vapor	liquid
Diameter	(ft)	0.1640	0.1640	0.1640
Elevation (Base)	(ft)	0.0000	0.0000	0.0000
Elevation (Ground)	(ft)	0.0000	0.0000	0.0000
Elevation (% of Height)	(%)	---	---	---
Dynamic Tank Factor				

Level Taps: Level Tap Specification

Level Tap	PV High	PV Low	OP High	OP Low

Level Taps: Calculated Level Tap Values

Level Tap	Liquid Level	Aqueous Level

Options

PV Work Term Contribution	(%)	100.00	

CONDITIONS

Name		Feed	liquid	vapor
Vapour		0.4511	0.0000	1.0000
Temperature	(F)	90.0000 *	90.0000	90.0000
Pressure	(psia)	370.0000 *	370.0000	370.0000
Molar Flow	(lbmole/hr)	10000.0000 *	5488.7051	4511.2949
Mass Flow	(lb/hr)	385619.0179	258133.7813	127485.2366
Std Ideal Liq Vol Flow	(barrel/day)	56003.9106	34142.7982	21861.1123
Molar Enthalpy	(Btu/lbmole)	-4.711e+004	-5.458e+004	-3.802e+004
Molar Entropy	(Btu/lbmole-F)	30.55	24.32	38.13
Heat Flow	(Btu/hr)	-4.7110e+08	-2.9956e+08	-1.7154e+08

PROPERTIES

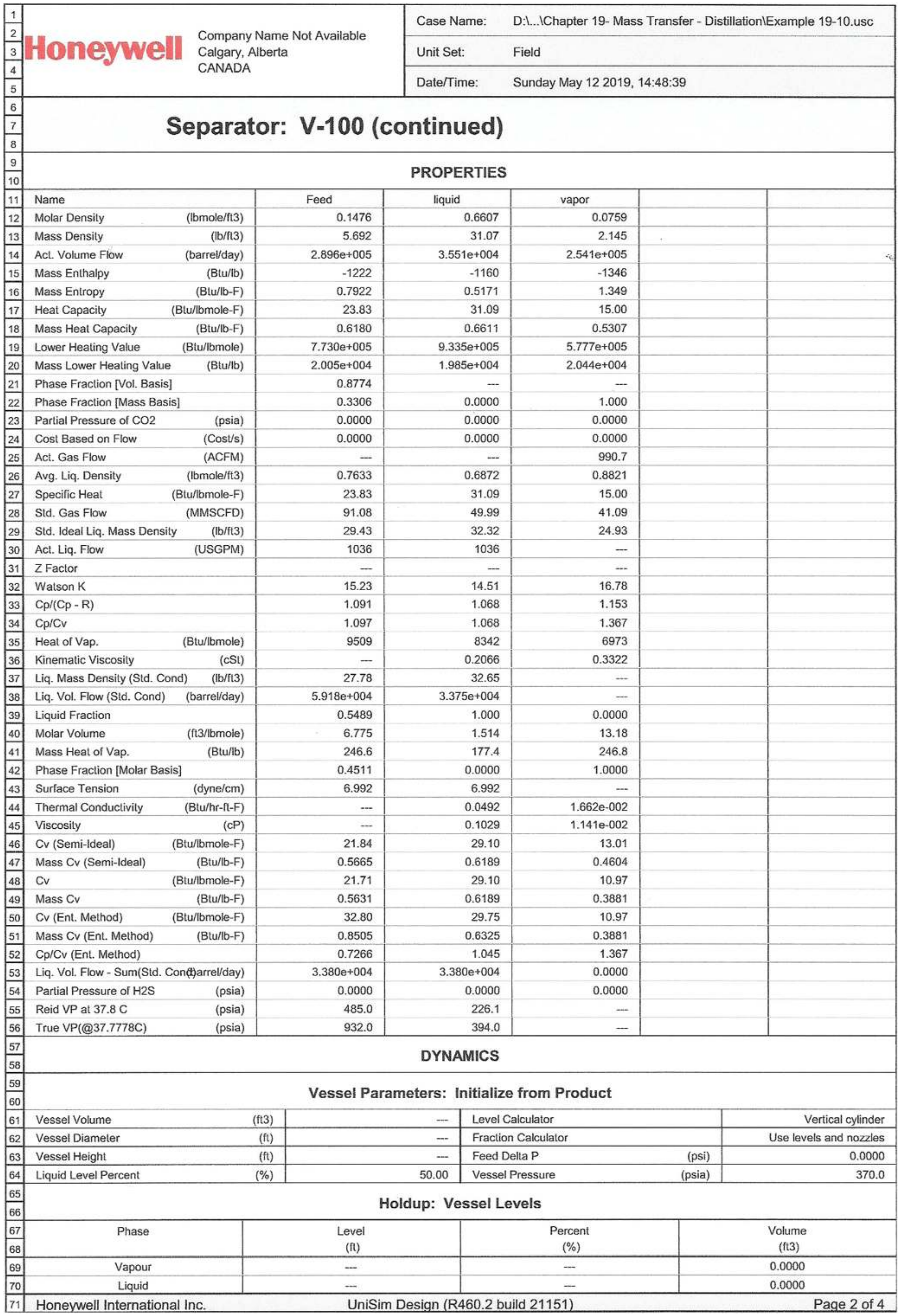
Name	Feed	liquid	vapor
Molecular Weight	38.56	47.03	28.26

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 1 of 4

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

(*Continued*)

Table 19.21 Results of the simulation using UniSim® Design R460.2 Software for Example 19.10. (*Continued*)



Honeywell Company Name Not Available
Calgary, Alberta
CANADA

Case Name: D:\...\Chapter 19- Mass Transfer - Distillation\Example 19-10.usc
Unit Set: Field
Date/Time: Sunday May 12 2019, 14:48:39

Separator: V-100 (continued)

PROPERTIES

Name		Feed	liquid	vapor		
Molar Density	(lbmole/ft3)	0.1476	0.6607	0.0759		
Mass Density	(lb/ft3)	5.692	31.07	2.145		
Act. Volume Flow	(barrel/day)	2.896e+005	3.551e+004	2.541e+005		
Mass Enthalpy	(Btu/lb)	-1222	-1160	-1346		
Mass Entropy	(Btu/lb-F)	0.7922	0.5171	1.349		
Heat Capacity	(Btu/lbmole-F)	23.83	31.09	15.00		
Mass Heat Capacity	(Btu/lb-F)	0.6180	0.6611	0.5307		
Lower Heating Value	(Btu/lbmole)	7.730e+005	9.335e+005	5.777e+005		
Mass Lower Heating Value	(Btu/lb)	2.005e+004	1.985e+004	2.044e+004		
Phase Fraction [Vol. Basis]		0.8774	---	---		
Phase Fraction [Mass Basis]		0.3306	0.0000	1.000		
Partial Pressure of CO2	(psia)	0.0000	0.0000	0.0000		
Cost Based on Flow	(Cost/s)	0.0000	0.0000	0.0000		
Act. Gas Flow	(ACFM)	---	---	990.7		
Avg. Liq. Density	(lbmole/ft3)	0.7633	0.6872	0.8821		
Specific Heat	(Btu/lbmole-F)	23.83	31.09	15.00		
Std. Gas Flow	(MMSCFD)	91.08	49.99	41.09		
Std. Ideal Liq. Mass Density	(lb/ft3)	29.43	32.32	24.93		
Act. Liq. Flow	(USGPM)	1036	1036	---		
Z Factor		---	---	---		
Watson K		15.23	14.51	16.78		
Cp/(Cp - R)		1.091	1.068	1.153		
Cp/Cv		1.097	1.068	1.367		
Heat of Vap.	(Btu/lbmole)	9509	8342	6973		
Kinematic Viscosity	(cSt)	---	0.2066	0.3322		
Liq. Mass Density (Std. Cond)	(lb/ft3)	27.78	32.65	---		
Liq. Vol. Flow (Std. Cond)	(barrel/day)	5.918e+004	3.375e+004	---		
Liquid Fraction		0.5489	1.000	0.0000		
Molar Volume	(ft3/lbmole)	6.775	1.514	13.18		
Mass Heat of Vap.	(Btu/lb)	246.6	177.4	246.8		
Phase Fraction [Molar Basis]		0.4511	0.0000	1.0000		
Surface Tension	(dyne/cm)	6.992	6.992	---		
Thermal Conductivity	(Btu/hr-ft-F)	---	0.0492	1.662e-002		
Viscosity	(cP)	---	0.1029	1.141e-002		
Cv (Semi-Ideal)	(Btu/lbmole-F)	21.84	29.10	13.01		
Mass Cv (Semi-Ideal)	(Btu/lb-F)	0.5665	0.6189	0.4604		
Cv	(Btu/lbmole-F)	21.71	29.10	10.97		
Mass Cv	(Btu/lb-F)	0.5631	0.6189	0.3881		
Cv (Ent. Method)	(Btu/lbmole-F)	32.80	29.75	10.97		
Mass Cv (Ent. Method)	(Btu/lb-F)	0.8505	0.6325	0.3881		
Cp/Cv (Ent. Method)		0.7266	1.045	1.367		
Liq. Vol. Flow - Sum(Std. Cond)	(barrel/day)	3.380e+004	3.380e+004	0.0000		
Partial Pressure of H2S	(psia)	0.0000	0.0000	0.0000		
Reid VP at 37.8 C	(psia)	485.0	226.1	---		
True VP(@37.7778C)	(psia)	932.0	394.0	---		

DYNAMICS

Vessel Parameters: Initialize from Product

Vessel Volume	(ft3)	---	Level Calculator		Vertical cylinder
Vessel Diameter	(ft)	---	Fraction Calculator		Use levels and nozzles
Vessel Height	(ft)	---	Feed Delta P	(psi)	0.0000
Liquid Level Percent	(%)	50.00	Vessel Pressure	(psia)	370.0

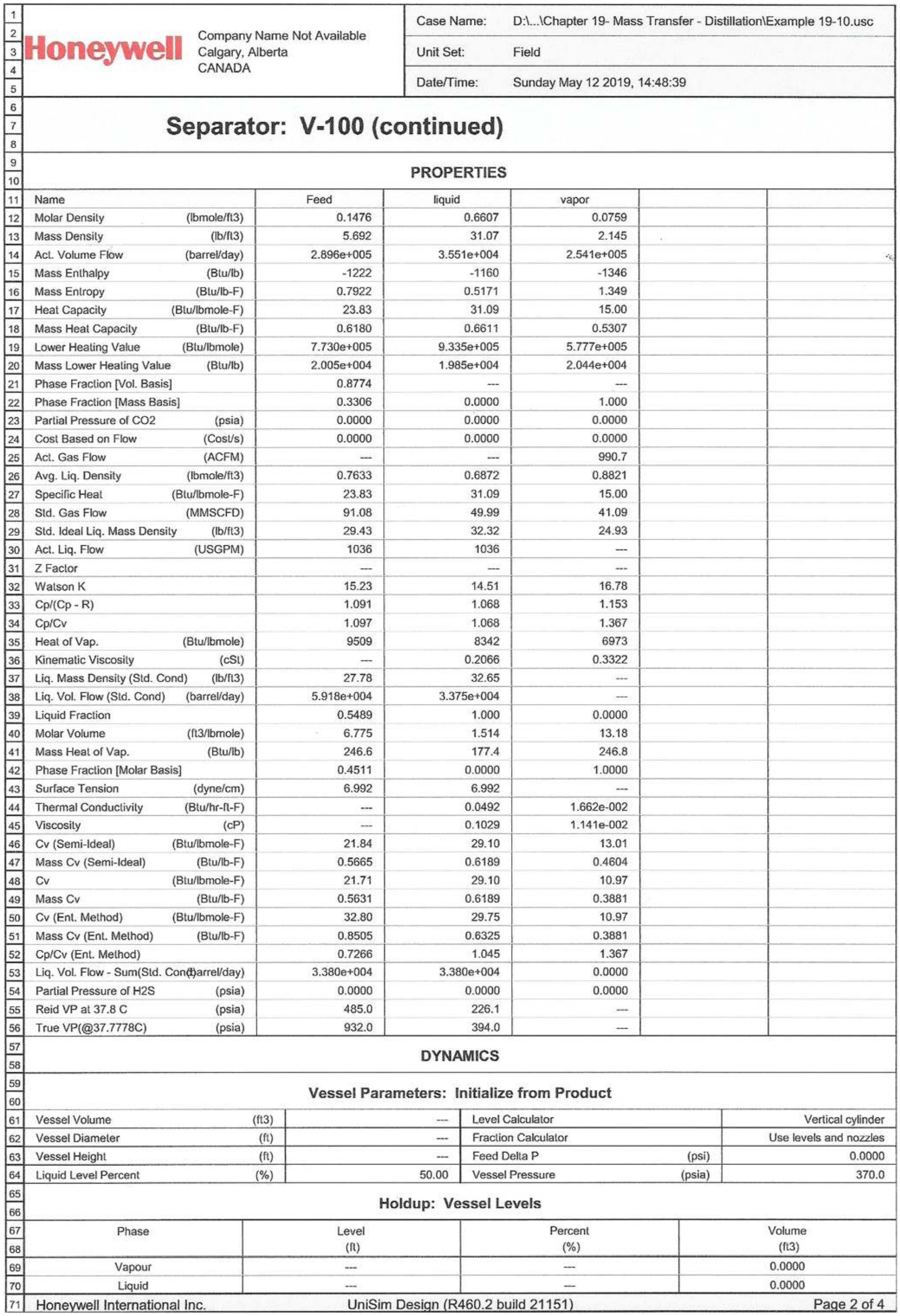
Holdup: Vessel Levels

Phase	Level (ft)	Percent (%)	Volume (ft3)
Vapour	---	---	0.0000
Liquid	---	---	0.0000

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 2 of 4

(*Continued*)

Table 19.21 Results of the simulation using UniSim® Design R460.2 Software for Example 19.10. (*Continued*)



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	D:\...\Chapter 19- Mass Transfer - Distillation\Example 19-10.usc
Unit Set:	Field
Date/Time:	Sunday May 12 2019, 14:48:39

Separator: V-100 (continued)

PROPERTIES

Name	Feed	liquid	vapor		
Molar Density (lbmole/ft3)	0.1476	0.6607	0.0759		
Mass Density (lb/ft3)	5.692	31.07	2.145		
Act. Volume Flow (barrel/day)	2.896e+005	3.551e+004	2.541e+005		
Mass Enthalpy (Btu/lb)	-1222	-1160	-1346		
Mass Entropy (Btu/lb-F)	0.7922	0.5171	1.349		
Heat Capacity (Btu/lbmole-F)	23.83	31.09	15.00		
Mass Heat Capacity (Btu/lb-F)	0.6180	0.6611	0.5307		
Lower Heating Value (Btu/lbmole)	7.730e+005	9.335e+005	5.777e+005		
Mass Lower Heating Value (Btu/lb)	2.005e+004	1.985e+004	2.044e+004		
Phase Fraction [Vol. Basis]	0.8774	---	---		
Phase Fraction [Mass Basis]	0.3306	0.0000	1.000		
Partial Pressure of CO2 (psia)	0.0000	0.0000	0.0000		
Cost Based on Flow (Cost/s)	0.0000	0.0000	0.0000		
Act. Gas Flow (ACFM)	---	---	990.7		
Avg. Liq. Density (lbmole/ft3)	0.7633	0.6872	0.8821		
Specific Heat (Btu/lbmole-F)	23.83	31.09	15.00		
Std. Gas Flow (MMSCFD)	91.08	49.99	41.09		
Std. Ideal Liq. Mass Density (lb/ft3)	29.43	32.32	24.93		
Act. Liq. Flow (USGPM)	1036	1036	---		
Z Factor	---	---	---		
Watson K	15.23	14.51	16.78		
Cp/(Cp - R)	1.091	1.068	1.153		
Cp/Cv	1.097	1.068	1.367		
Heat of Vap. (Btu/lbmole)	9509	8342	6973		
Kinematic Viscosity (cSt)	---	0.2066	0.3322		
Liq. Mass Density (Std. Cond) (lb/ft3)	27.78	32.65	---		
Liq. Vol. Flow (Std. Cond) (barrel/day)	5.918e+004	3.375e+004	---		
Liquid Fraction	0.5489	1.000	0.0000		
Molar Volume (ft3/lbmole)	6.775	1.514	13.18		
Mass Heat of Vap. (Btu/lb)	246.6	177.4	246.8		
Phase Fraction [Molar Basis]	0.4511	0.0000	1.0000		
Surface Tension (dyne/cm)	6.992	6.992	---		
Thermal Conductivity (Btu/hr-ft-F)	---	0.0492	1.662e-002		
Viscosity (cP)	---	0.1029	1.141e-002		
Cv (Semi-Ideal) (Btu/lbmole-F)	21.84	29.10	13.01		
Mass Cv (Semi-Ideal) (Btu/lb-F)	0.5665	0.6189	0.4604		
Cv (Btu/lbmole-F)	21.71	29.10	10.97		
Mass Cv (Btu/lb-F)	0.5631	0.6189	0.3881		
Cv (Ent. Method) (Btu/lbmole-F)	32.80	29.75	10.97		
Mass Cv (Ent. Method) (Btu/lb-F)	0.8505	0.6325	0.3881		
Cp/Cv (Ent. Method)	0.7266	1.045	1.367		
Liq. Vol. Flow - Sum(Std. Cond) (barrel/day)	3.380e+004	3.380e+004	0.0000		
Partial Pressure of H2S (psia)	0.0000	0.0000	0.0000		
Reid VP at 37.8 C (psia)	485.0	226.1	---		
True VP(@37.7778C) (psia)	932.0	394.0	---		

DYNAMICS

Vessel Parameters: Initialize from Product

Vessel Volume	(ft3)	---	Level Calculator		Vertical cylinder
Vessel Diameter	(ft)	---	Fraction Calculator		Use levels and nozzles
Vessel Height	(ft)	---	Feed Delta P	(psi)	0.0000
Liquid Level Percent	(%)	50.00	Vessel Pressure	(psia)	370.0

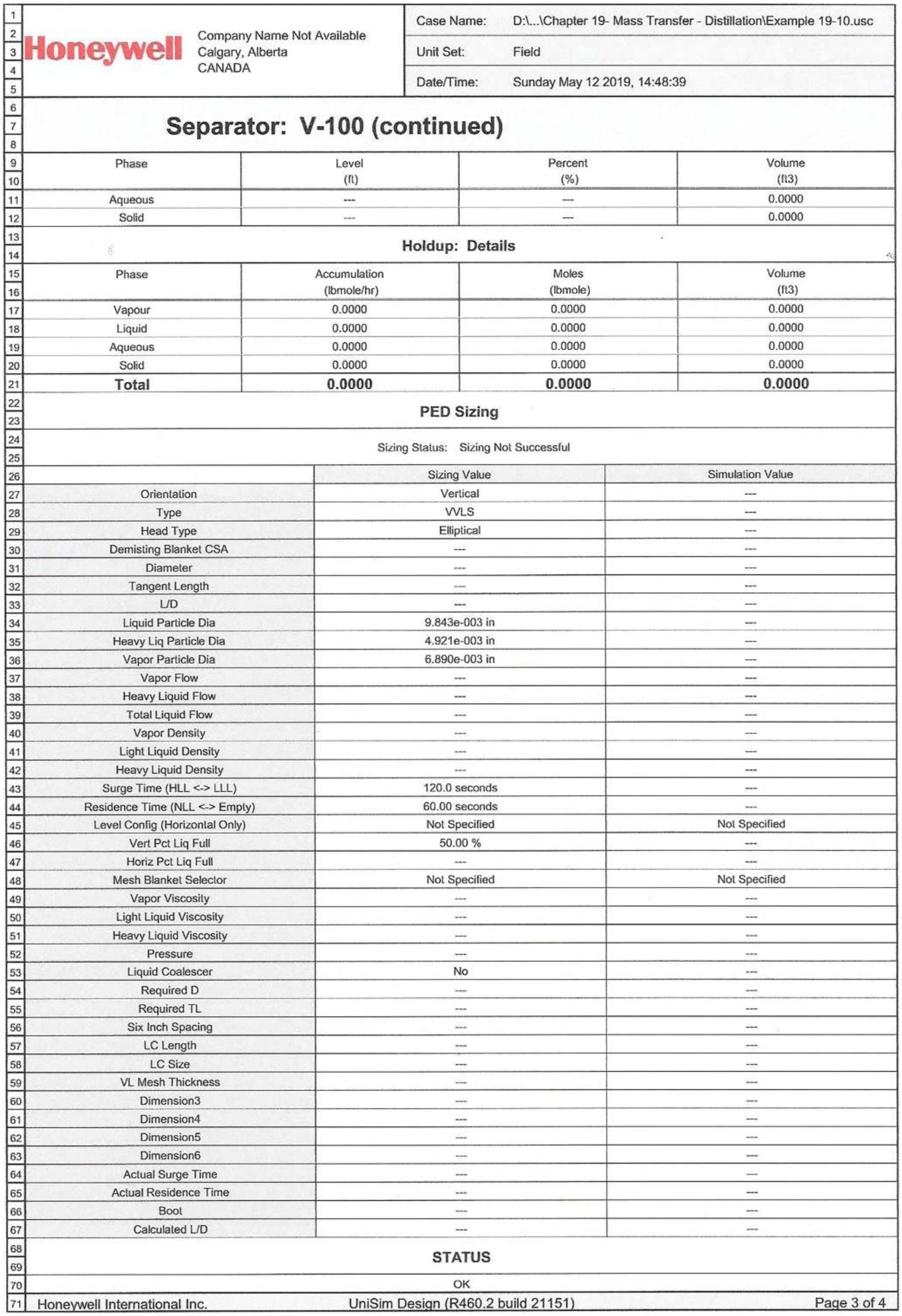
Holdup: Vessel Levels

Phase	Level (ft)	Percent (%)	Volume (ft3)
Vapour	---	---	0.0000
Liquid	---	---	0.0000

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 2 of 4

(*Continued*)

Table 19.21 Results of the simulation using UniSim® Design R460.2 Software for Example 19.10. (*Continued*)



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name: D:\...\Chapter 19- Mass Transfer - Distillation\Example 19-10.usc
Unit Set: Field
Date/Time: Sunday May 12 2019, 14:48:39

Separator: V-100 (continued)

Phase	Level (ft)	Percent (%)	Volume (ft3)
Aqueous	---	---	0.0000
Solid	---	---	0.0000

Holdup: Details

Phase	Accumulation (lbmole/hr)	Moles (lbmole)	Volume (ft3)
Vapour	0.0000	0.0000	0.0000
Liquid	0.0000	0.0000	0.0000
Aqueous	0.0000	0.0000	0.0000
Solid	0.0000	0.0000	0.0000
Total	**0.0000**	**0.0000**	**0.0000**

PED Sizing

Sizing Status: Sizing Not Successful

	Sizing Value	Simulation Value
Orientation	Vertical	---
Type	VVLS	---
Head Type	Elliptical	---
Demisting Blanket CSA	---	---
Diameter	---	---
Tangent Length	---	---
L/D	---	---
Liquid Particle Dia	9.843e-003 in	---
Heavy Liq Particle Dia	4.921e-003 in	---
Vapor Particle Dia	6.890e-003 in	---
Vapor Flow	---	---
Heavy Liquid Flow	---	---
Total Liquid Flow	---	---
Vapor Density	---	---
Light Liquid Density	---	---
Heavy Liquid Density	---	---
Surge Time (HLL <-> LLL)	120.0 seconds	---
Residence Time (NLL <-> Empty)	60.00 seconds	---
Level Config (Horizontal Only)	Not Specified	Not Specified
Vert Pct Liq Full	50.00 %	---
Horiz Pct Liq Full	---	---
Mesh Blanket Selector	Not Specified	Not Specified
Vapor Viscosity	---	---
Light Liquid Viscosity	---	---
Heavy Liquid Viscosity	---	---
Pressure	---	---
Liquid Coalescer	No	---
Required D	---	---
Required TL	---	---
Six Inch Spacing	---	---
LC Length	---	---
LC Size	---	---
VL Mesh Thickness	---	---
Dimension3	---	---
Dimension4	---	---
Dimension5	---	---
Dimension6	---	---
Actual Surge Time	---	---
Actual Residence Time	---	---
Boot	---	---
Calculated L/D	---	---

STATUS

OK

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 3 of 4

(*Continued*)

Table 19.21 Results of the simulation using UniSim® Design R460.2 Software for Example 19.10. (*Continued*)

Honeywell
Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	D:\...\Chapter 19 Mass Transfer - Distillation\Example 19-10.usc
Unit Set:	Field
Date/Time:	Sunday May 12 2019, 14:48:39

Separator: V-100 (continued)

NOTES

Description

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 4 of 4

19.18 Binary System Material Balance: Constant Molal Overflow Tray to Tray

Refer to Figure 19.5. (For an overall review, see Reference [65])

- Rectifying Section

$$V_r = L_r + D \tag{19.132}$$

For any component in the mixture; using total condenser see Figures 19.5 and 19.33.

$$V_n y_{ni} = L_{n+1}\, x_{(n+1)i} + D x_{Di} \tag{19.133}$$

$$y_{ni} = \frac{L_{n+1}}{V_n} x_{(n+1)i} + \frac{D}{V_n} x_{Di} \tag{19.134}$$

- Operating Line Equation:

$$y_{ni} = \frac{L_r}{V_r} x_{(n+1)i} + \frac{D}{V_r} x_{Di} \tag{19.135}$$

For total condenser: y (top plate) = x_D

- Stripping Section:

$$L_s = V_s + B \tag{19.136}$$

$$L_{(m+1)}\, x_{(m+1)i} = V_m\, y_{mi} + Bx_{Bi} \tag{19.137}$$

- Operating Line Equation:

$$y_{mi} = \frac{L_s}{V_s} x_{(m+1)i} - \frac{B}{V_s} x_{Bi} \tag{19.138}$$

19.18.1 Conditions of Operation (Usually Fixed)

1. Feed composition, and quantity.
2. Reflux Ratio (this is one of the initial unknowns).
3. Thermal condition of feed (e.g., liquid at boiling point, all vapor, sub-cooled liquid).
4. Degree, type or amount of fractionation or separation, including compositions of overhead or bottoms.
5. Column operating pressure or temperature of condensation of overhead (determined by temperature of cooling medium), including type of condensation, i.e., total or partial.
6. Constant molal overflow from stage to stage (theoretical) for simple ideal systems following Raoult's Law. Techniques that are more complicated apply for non-ideal systems.

19.18.2 Flash Vaporization

At a total pressure, P, the temperature of flash must be between the dew point and the bubble point of a mixture [66–70] (see Figure 19.31). Thus:

T (Bubble Point) < T (Flash) < T (Dew Point)

For binary mixtures [69]:

1. Set the temperature and pressure of the flash chamber.
2. Perform a material balance on a single component.

$$F_t X_i = V_t y_i + L x_i \tag{19.139}$$

3. Knowing F, calculate amounts and compositions of V and L.

$F_t = F + V_s$ = mols of feed plus mols of non-condensable gases

From Henry's Law:

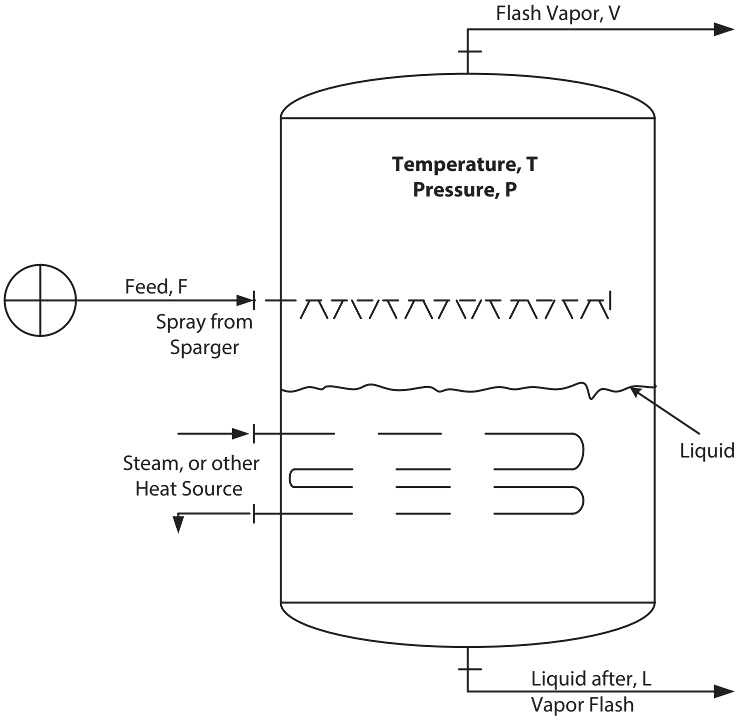


Figure 19.31 Schematic liquid flash tank. Note: Feed can be preheated to vaporize feed partially.

$$F_t X_{fi} = V_t y_i + L(y_i/K_i) \tag{19.140}$$

$V_t = V + V_s$ = mols of vapor formed plus mols non-condensed gases

$$y_i = K_i x_i$$

4.

$$F_t = V_t + L \tag{19.141}$$

$$y_i = \frac{F_t X_{fi}}{V_t + \frac{L}{K_i}} = \frac{F_t X_{fi}}{Ft - L + \frac{L}{K_i}} \tag{19.142}$$

$$y_i = \frac{X_{fi}}{1 - \frac{L}{F_t}\left(1 - \frac{1}{K_i}\right)} \tag{19.143}$$

After calculating V, calculate the x_i's and y_i's:

$$y_i V = \frac{FX_i}{1 + \frac{L}{K_i V}} \tag{19.144}$$

Then,

$$y_i = \frac{\left(\frac{FX_i}{V}\right)}{1+\frac{L}{K_iV}} \tag{19.145}$$

Calculate each y; after calculating the y_i's, calculate x_i's as follows:

$$y_i = K_i x_i \tag{19.38}$$

$$\text{so, } x_i = y_i/K_i$$

$$K_i = P_i/\pi \tag{19.146}$$

Relation for K:

$$y_i = K_i x_i \tag{19.38}$$

$$p_i = P_i x_i \tag{19.147}$$

$$p_i = \pi y_i \tag{19.148}$$

$$\pi y_i = P_i x_i \tag{19.149}$$

$$\frac{y_i}{x_i} = \frac{P_i}{\pi} \tag{19.150}$$

where

X_i = mols of a component i in vapor plus mols of same component in liquid, divided by total mols of feed (both liquid and vapor)
= total mol fraction, irrespective of whether component is in liquid or vapor.

$$FX_i = Vy_i + Lx_i \tag{19.151}$$

$$x_i = y_i/K_i$$

$$y_i = V + L/K_i \tag{19.152}$$

$$y_i V = V\left(1+\frac{L}{K_iV}\right) \tag{19.153}$$

$$\Sigma y_i = 1.0$$

$$\sum_{i=1}^{c} y_i V = \sum_{i=1}^{c}\left(\frac{FX_i}{1+\frac{L}{K_i V}}\right) \tag{19.154}$$

To calculate,
V, L, y_i’ s, and x_i’s:

1. Assume: V
2. Calculate: L = F - V
3. Calculate: L/V
4. Look up K_i’s at temperature and total pressure of system
5. Substitute in:

$$V = \sum_{i=1}^{c} \frac{FX_i}{1+\frac{L}{K_i V}} \tag{19.155}$$

Composition	**Mol Fraction**	**Assumed T = 90°F K**	**Kx**	**Assume T = 100°F, K**	**Kx**
C_2H_6	0.15	3.1	0.465	3.4	0.5100
C_3H_8	0.15	1.0	0.150	1.2	0.1800
n-C_4H_{10}	0.30	0.35	0.105	0.39	0.1170
i-C_4H_{10}	0.25	0.46	0.115	0.52	0.1300
n-C_5H_{12}	0.15	0.12	0.018	0.13	0.0195
	1.00		0.853		0.9565
					(Too low)
		Assume T = 105°F			
			K	Kx	
			3.45	0.5175	
			1.25	0.1875	
			0.41	0.1230	
			0.55	0.1375	
			0.15	0.0225	
				0.9880	

6. If an equality is obtained from:

$V_{calc} = V_{assumed}$, then the amount of vapor was satisfactory as assumed.

$$V = \frac{FX_1}{1+\frac{L}{K_1V}} + \frac{FX_2}{1+\frac{L}{K_2V}} + \frac{FX_3}{1+\frac{L}{K_3V}} + \ldots \frac{FX_i}{1+\frac{L}{K_iV}} \qquad (19.156)$$

7.

$$Vy_i = \frac{FX_i}{1+\frac{L}{K_iV}} \qquad (19.144)$$

$$\text{and } y_i = \frac{FX_i/V}{1+\frac{L}{K_iV}} \qquad (19.145)$$

8. Calculate each y_i as in (7) above, then the x_i's are determined:

$$y_i = K_i x_i$$
$$\text{or, } x_i = y_i/K_i$$

9.

$$K_i = P_i/\pi \qquad (19.146)$$

where

π = total system pressure absolute.
P_i = vapor pressure of individual component at temperature, abs.
$X_i = X$ = mols of a component i, in vapor phase plus mols of same component in liquid divided by the total mols of feed (both liquid and vapor).
x_f = mol fraction of a component in feed.
x_f = mol fraction of any component in the feed, F_t where $X_f = Fx_f/F_t$ for all components in F; for the non-condensable gases, $x_f = V_s/F_t$.
F = mols of feed entering flash zone per unit time contains all components except non-condensable gases.
F_t = $F + V_s$
V_t = $V + V_s$ mols of vapor at a specific temperature and pressure, leaving flash zone per unit time.
V_s = mols of non-condensable gases entering with the feed, F, and leaving with the vapor, V, per unit time.
V = mols of vapor produced from F per unit time
F = V + L
L = mols of liquid at a specific temperature and pressure, from F, per unit time.
i = specific individual component in mixture.
K_i = equilibrium K values for a specific component at a specific temperature and pressure, from References [28], [49], and [71–74].

T = temperature, abs.

x_i = mol fraction of a specific component in liquid mixture as may be associated with feed, distillate, or bottoms, respectively.

y_i = mol fraction of a specific component in vapor mixture as may be associated with the feed, distillate or bottoms, respectively.

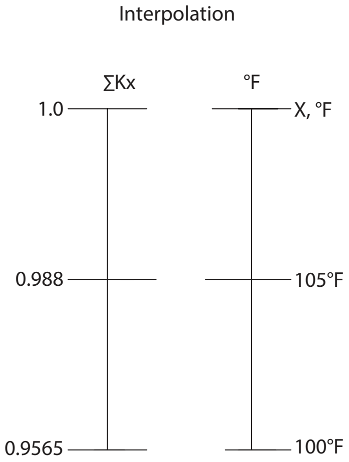
10. For the simplified case of a mixture free of non-condensable gases, see Equation 10.143, where $X_{fi} = x_f$.

Example 19.11: Bubble Point and Dew Point

From the hydrocarbon feed stock listed, calculate the bubble point and dew point of the mixture at 165 psia, and using K values as listed, which can be read from a chart in 3rd edition Perry's, Chemical Engineer's Handbook.
Feed Stock:

Composition	**Mol Fraction**
C_2H_6	0.15
C_3H_8	0.15
n-C_4H_{10}	0.30
i-C_4H_{10}	0.25
n-C_4H_{12}	0.15
	1.00

Calculate the Bubble Point (Assume composition is liquid)



By interpolation:

By interpolation:

$$\frac{105-100}{0.988-0.9565}=\frac{X-100}{1.0-0.9565}$$

$$X = 106.9°F\ (107°F)$$

So, T = 109°F Bubble point at 165 psia

Calculation of Dew Point

Composition in vapor	Mol. Frac. in vapor	Assume, T = 160°F K (from charts)	x = y/K
C_2H_6	0.15	5.1	0.0294
C_3H_8	0.15	1.85	0.0811
n-C_4H_{10}	0.30	0.80	0.3750
i-C_4H_{10}	0.25	1.00	0.2500
n-C_5H_{12}	0.15	0.32	0.4688
	1.00		1.204 ≠ Σy/K = 1

Composition in vapor	Mol. Frac. in vapor	Assume, T = 180°F K (from charts)	x = y/K
C_2H_6	0.15	5.95	0.0252
C_3H_8	0.15	2.25	0.0666
n-C_4H_{10}	0.30	0.98	0.3061
i-C_4H_{10}	0.25	1.34	0.1866
n-C_5H_{12}	0.15	0.41	0.3659
	1.00		0.9504 ≠ Σy/K = 1

Composition in vapor	Mol. Frac. in vapor	Assume, T = 175°F K (from charts)	x = y/K
C_2H_6	0.15	5.60	0.0268
C_3H_8	0.15	2.20	0.0682
n-C_4H_{10}	0.30	0.91	0.3297
i-C_4H_{10}	0.25	1.20	0.2083
n-C_5H_{12}	0.15	0.39	0.3846
	1.00		1.0176 ≅ Σy/K = 1

Dew point is essentially 175°F at 165 psia

Example 19.12: Calculation of Flashing Composition

If the mixture shown in Example 19.11 is flashed at a temperature midway between the bubble point and dew point, and at a pressure of 75 psia, calculate the amounts and compositions of the gas and liquid phases.

Referring to Example 19.10:

Composition	Mol Fraction
C_2H_6	0.15
C_3H_8	0.15
n-C_4H_{10}	0.30
i-C_4H_{10}	0.25
n-C_5H_{12}	0.15
	1.00

Must Calculate Bubble Point at 75 psia

Composition	Mol Frac.	K @ 50°F	y = Kx	K @ 40°F	y = Kx
C_2H_6	0.15	5.0	0.7500	4.5	0.6750
C_3H_8	0.15	1.2	0.1800	1.07	0.1605
n-C_4H_{10}	0.30	0.325	0.0975	0.28	0.0840
i-C_4H_{10}	0.25	0.48	0.1200	0.415	0.10380
n-C_5H_{12}	0.15	0.089	0.0134	0.074	0.0110
			1.1609		1.0343

Bubble point = 40°F (as close as K curves can be read)

$$\text{Extrapolating} = \left(\frac{1.00-1.0343}{1.1609-1.0343}\right)(50^\circ-40^\circ)=2.7^\circ$$

Therefore, a close value of bubble point would be: 40° − 3° = 37°F

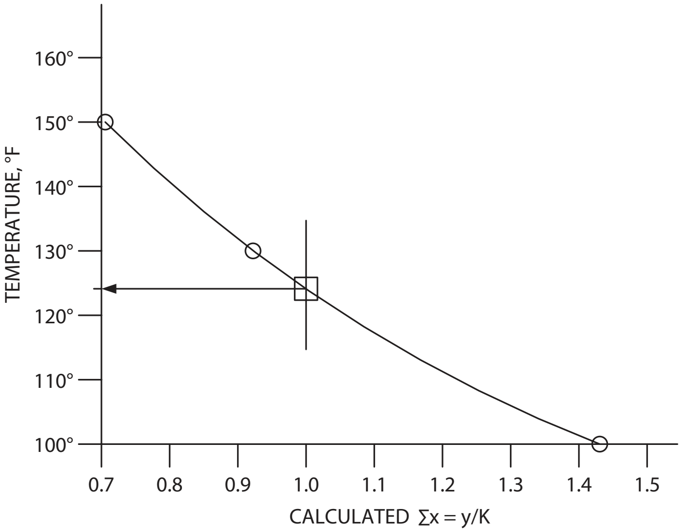


Figure 19.32 Extrapolation curve for dew point for Example 19.12.

Calculate Dew Point at 75 psia

Composition	Mol Frac.	K @ 70°F	Σx = y/K	K @ 100°F	Σx = y/K	K @ 130°F	Σx = y/K
C_2H_6	0.15	5.9	0.0254	7.6	0.0197	9.5	0.0158
C_3H_8	0.15	1.55	0.0968	2.18	0.0688	3.0	0.0500
n-C_4H_{10}	0.30	0.45	0.6667	0.70	0.4286	1.06	0.2830
i-C_4H_{10}	0.25	0.58	0.4310	1.0	0.2500	1.45	0.1724
n-C_5H_{12}	0.15	0.13	1.1538	0.225	0.6667	0.37	0.4054
			2.3737		1.4338		0.9266
		(too low temp.)					

Refer to extrapolation curve, Figure 19.32.

At Σx = y/K = 1.0, dew point = 124°F

Flash this mixture at temperature midway between bubble point and dew point, or flash temperature = $\frac{37+124}{2}$ = 81°F

Assume: F (feed) = 100

Pressure: = 75 psia; then tabulating the calculations:

Composition	Feed Mol. Frac., x	FX	L/V	K@81°F	$\frac{L}{VK}$	$1+\frac{L}{VK}$	$\sum yV = \frac{FX}{1+\frac{L}{VK}}$
C_2H_6	0.15	15		6.6	0.3545	1.3545	11.0742
C_3H_8	0.15	15	2.34	1.78	1.3146	2.3146	6.4806
n-C_4H_{10}	0.30	30		0.54	4.3333	5.3333	5.6250
i-C_4H_{10}	0.25	25		0.77	3.0390	4.0390	6.1896
n-C_5H_{12}	0.15	15		0.16	14.6250	15.6250	0.9600
							30.33

Vapor phase after flashing at 75 psia and 81°F = 30.3% of original feed.

Liquid phase = 100 − 30 = 70% of original feed

Composition of Vapor

Composition	Mol Fraction y
C_2H_6	11.074/30.33 = 0.3651
C_3H_8	6.481/30.33 = 0.2137
n-C_4H_{10}	5.625/30.33 = 0.1855
i-C_4H_{10}	6.19/30.33 = 0.2041
n-C_5H_{12}	0.96/30.33 = 0.0317
	$\sum y = 1.000$

Composition of Liquid

Feed Composition	K @ 81°F	Mol Fraction x = y/K
C_2H_6	6.6	0.3651/6.6 = 0.0553
C_3H_8	1.78	0.2137/1.78 = 0.1201
n-C_4H_{10}	0.54	0.1855/0.54 = 0.3435
i-C_4H_{10}	0.77	0.2041/0.77 = 0.2651
n-C_5H_{12}	0.16	0.0317/0.16 = 0.1981
		$\sum x = 0.9821$

This should be = 1.00. Inaccuracy in reading K values probably accounts for most of the difference.

19.19 Determination of Distillation Operating Pressures

Determining the proper operating pressure for a distillation system, whether using trays or packed column, requires examining the conditions by following the pattern of Figure 19.33 [75]. It is essential to establish the condensing conditions of the distillation overhead vapors, and any limitations on bottoms temperature at the estimated pressure drop through the system. Preliminary calculations for the number of trays or amount of packing must be performed to develop a reasonable system pressure drop. With this accomplished, the top and bottom column conditions can be established, and more detailed calculations performed. For trays this can be 0.1 psi/actual tray to be installed [75] whether operating at atmospheric or above, and use 0.05 psi/tray equivalent for low vacuum (not low absolute pressure).

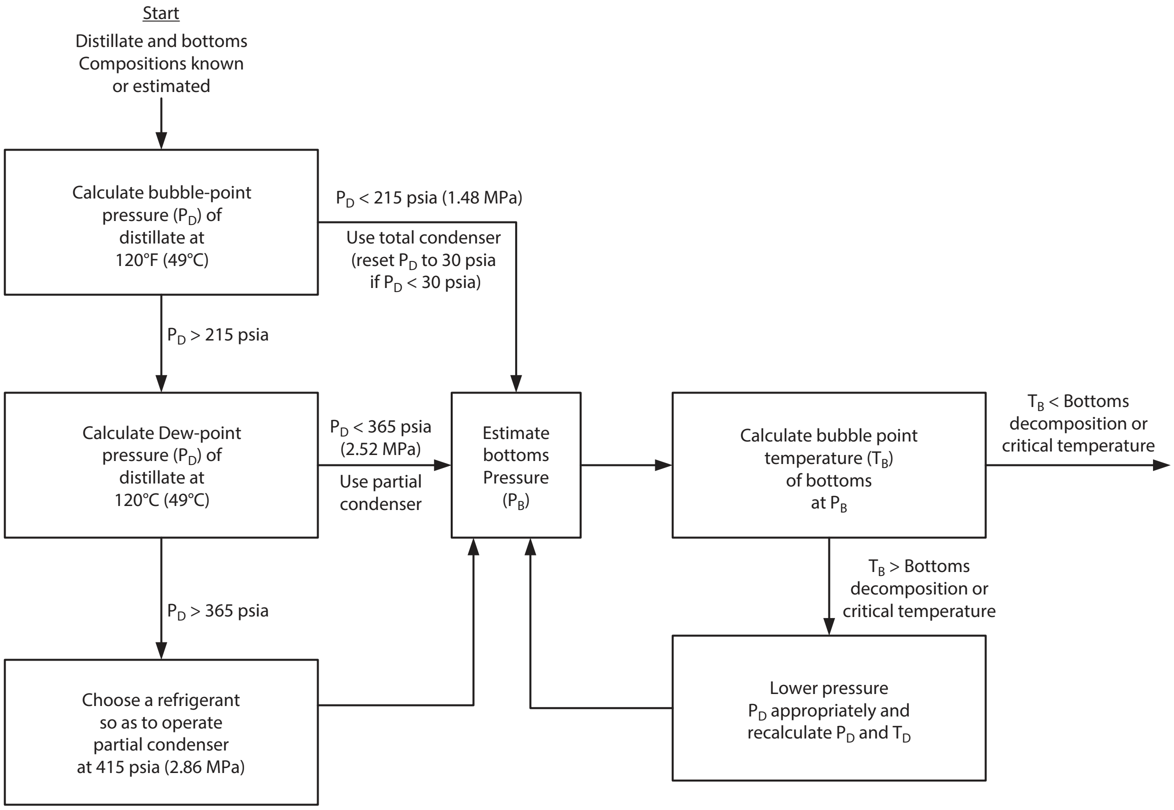


Figure 19.33 Algorithm for establishing distillation column pressure and condenser type (source: Henley, E. J. and Seader, J. D., Separation Process Principles, 2nd Edition, John Wiley © p265 [2006]).

Because low-pressure operations require larger diameter columns, selected pressures should only be as low as required to accomplish the separation.

For high vacuum distillation, Eckles *et al.* [76] suggest using a thin film or conventional batch process for industrial type installations. However, many tray and packed columns operate at pressures as low as 4 mm Hg. abs. Eckles [76] suggests "high vacuum" be taken as 5 mm Hg; molecular distillation be 0.3-0.003 mm Hg pressure, and unobstructed path distillation occur at 0.5-0.02 mm Hg. These latter two can be classed as evaporation processes. Eckles' [76] rules of thumb can be summarized:

1. Do not use a lower pressure than necessary, because separation efficiency and throughput decrease as pressure decreases.
2. The requirement to select a bottoms temperature to avoid overheating heat sensitive materials may become controlling.
3. When separating volatile components such as a single stream from low-volatility bottoms, use a molecular or unobstructed path process of, either thin film or batch type.
4. When separating a volatile product from volatile impurities, batch distillation is usually preferable.
5. Do not add a packed column to a thin film evaporator system, because complications may occur.

Note that good vapor–liquid equilibrium data for low pressure conditions are very scarce and difficult to locate. However, they are essential for proper calculations. This is dealt with in References [77] and [78].

Studies with high-pressure distillation by Brierley [79] provide insight into some FRI studies, the effects of pressure on performance, and the impacts of errors in physical properties, relative volatility, etc. This work makes an important contribution to understanding and setting of operating pressures.

19.20 Condenser Types From a Distillation Column

19.20.1 Total Condenser

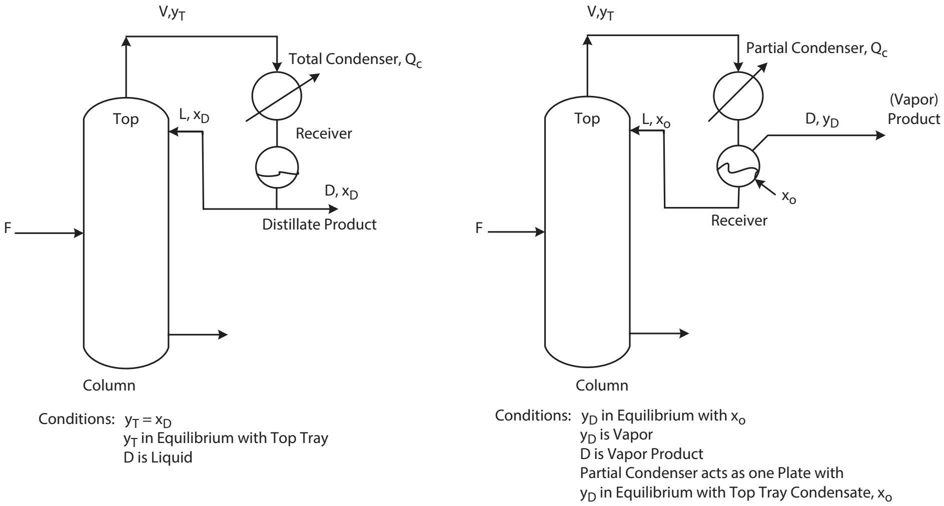
In a total condenser all the overhead vapor is condensed to the liquid. The heat load on the condenser is equal to the sum of the sensible heat above the dew point, plus the latent of vapor, plus the sensible heat in cooling liquid down to the bubble point. Condensation of liquid will commence at the dew point and end at the bubble point. (Any further removal would proceed due to cooling). The condenser and accumulator pressure will be the total vapor pressure of the condensate. If an inert gas is present, the system total pressure will be affected accordingly. When using a total condenser, the condensed stream is subsequently split into two streams. One is returned into the column as reflux and the second leaves as distillate product.

19.20.2 Partial Condenser

The effect of the partial condenser is indicated in Figure 19.34 and is represented by the relations given in the rectifying and stripping sections. The product is a vapor that is in equilibrium with the reflux to the column top tray, and hence the partial condenser actually serves as an "external" tray for the system; it should be treated as the top tray when applying the equations for total reflux conditions. This requires care in step-wise calculation of the column performance.

In a partial condenser there are two general conditions of operation:

1. All condensed liquid is returned to column as reflux, and all vapor is withdrawn from the accumulator as product. In this case the vapor $y_c = x_D$; Figure 19.5 and Figure 19.35.
2. Both liquid and vapor products are withdrawn, with liquid reflux composition being equal to liquid product composition. On an equilibrium diagram the partial condenser liquid and vapor stream's respective compositions are in equilibrium, but only when combined do they represent the intersection of the operating line with the 45° slope (Figure 19.35).



Total Condenser

(**a**)

Partial Condenser

(**b**)

Figure 19.34 Total and partial condenser arrangements.

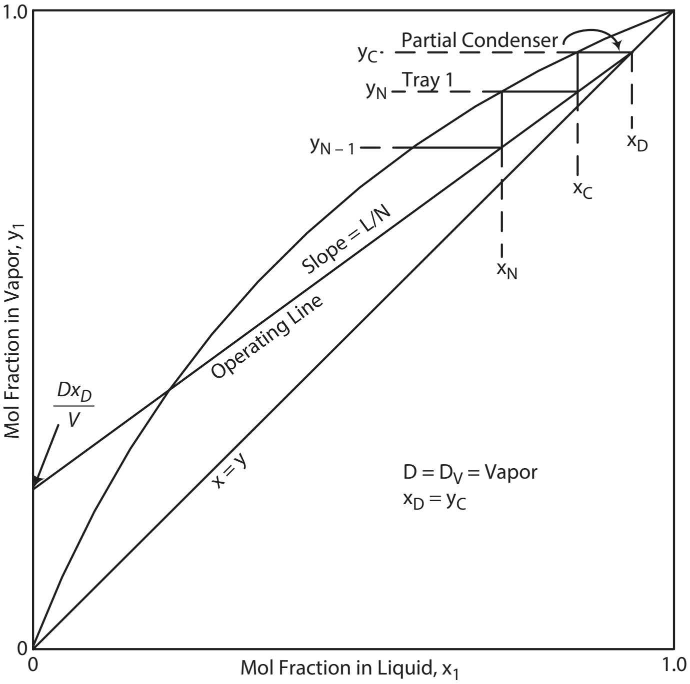


Figure 19.35 Diagram of partial condenser; only a vapor product is withdrawn.

Example 19.13: Calculation of Partial Condensation of a Multicomponent Feed

A multicomponent vapor feed (F) is partially condensed to produce a vapor (V) and a liquid (L). Mole fractions of components in the vapor feed are designated z, those in the vapor y, and those in the liquid x. The vapor-liquid equilibrium for any component is given by y = Kx. Assuming that the vapor and the liquid leave the partial condenser in equilibrium, show that for any component,

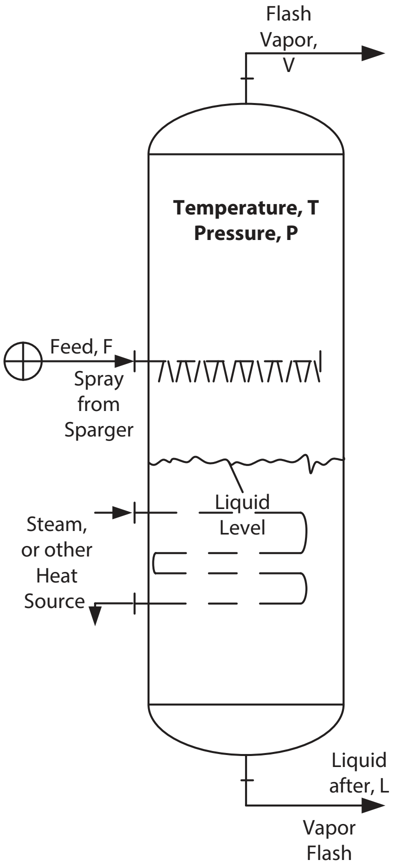
$$y = \frac{z(F/V)}{(1 + L/KV)}$$

It has been calculated that at 18.9 bar and 55°C, one third of the vapor feed will be condensed.

Check this calculation for the feed composition and K values at 18.9 bar and 55°C given as:

Component	C_2H_6	C_3H_8	C_4H_{10}
z	0.0713	0.8978	0.0309
K	2.65	1.0	0.41

Solution



Mass balance at the top of the column:

$$F = V + L \tag{19.157}$$

Component balance:

$$Fz = Vy + Lx \tag{19.158}$$

Since the vapor phase is in equilibrium with the liquid phase at the top of the column.

$$y = K\,x \tag{19.159}$$

$$x = y/K \tag{19.160}$$

Substitute Equation 19.160 into Equation 19.158 gives:

$$Fz = Vy + L\frac{y}{K} \tag{19.161}$$

or

$$Fz = y\left(V + \frac{L}{K}\right) \tag{19.162}$$

$$y = \frac{Fz}{V + \frac{L}{K}} = \frac{(F/V)z}{\left(1 + \frac{L}{KV}\right)} \tag{19.163}$$

Let F = 3 kmol/h

Since 1/3 of the vapor feed is condensed, then

$$L = 1/3 \text{ of } 3 \text{ kmol/h} = 1 \text{ kmol/h}$$
$$V = F - L = 2 \text{ kmol/h}$$

To check the total mole fraction of the components at the top of the column, we determine the mole fraction of individual component from Equation 19.163, i.e.,

Composition	**Feed Mol. Frac., z**	**Fz**	**Fz/V**	**K@ 55°F**	$\frac{L}{VK}$	$1+\frac{L}{VK}$	$y=\frac{Fz/V}{\left(1+\frac{L}{VK}\right)}$
C_2H_6	0.07135	0.2139	0.10695	2.65	0.1887	1.1887	0.08997
n-C_3H_8	0.8978	2.6934	1.3467	1.00	0.5000	1.5	0.8978
n-C_4H_{10}	0.0309	0.0927	0.04635	0.41	1.2195	2.219	0.0209
						$\sum y$	1.00867

The condensing temperature range in a total condenser is dependent on the dew and bubble points, cooling water limitations, fouling, and the pressure of the distillation column. The reflux temperature is also affected by the pressure control scheme and the presence of excess capacity for cooling. The control aspect example can be referred to when submerged tubes result from some pressure control schemes. The degree of sub cooling can be as much as 50 to 200°F.

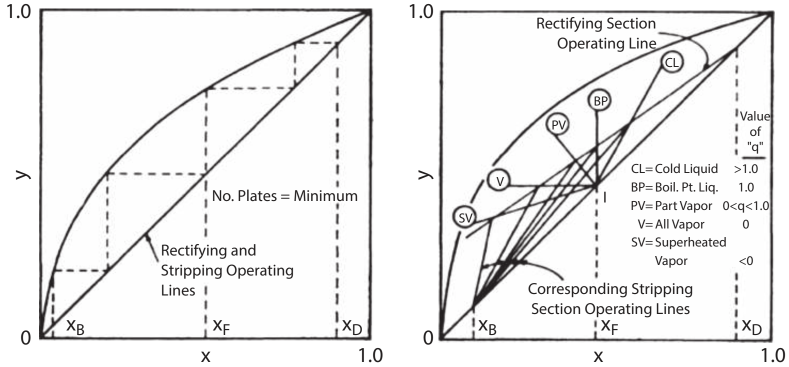
19.21 Effect of Thermal Condition of Feed

The condition of the feed as it enters the column has an effect on the number of trays, reflux requirements and heat energy requirements for a given separation. Figure 19.36 illustrates the possible situations, i.e., sub-cooled liquid feed, feed at the boiling point of the column feed tray, part vapor and part liquid, all vapor, or superheated vapor. The thermal condition is designated as "q"; this is approximately the amount of heat required to vaporize one mol of feed at the feed tray conditions, divided by the latent heat of vaporization of the feed.

For a single feed stream column with a total condenser and a partial reboiler as shown in Figure 19.5, a mass balance in the rectifying section of the more volatile component is:

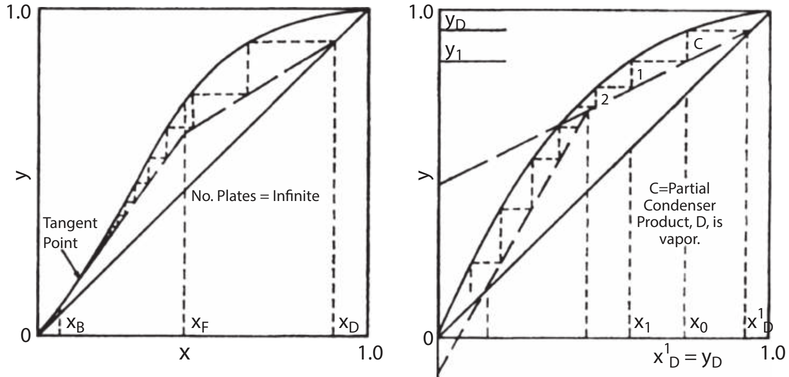
$$Vy = L\,x + D\,x_D \tag{19.164}$$

A mass balance in the stripping section is



Total reflux

Thermal condition of feed to column



Minimum reflux abnormal equilibrium

Partial condensation of overhead vapors

Figure 19.36 Operating characteristics of distillation columns.

$$\bar{V}y = \bar{L}x - Bx_B \tag{19.165}$$

Assuming constant molar overflow, at the feed plate the point at which the top operating line intersects the bottom operating line is:

$$x_{\text{top op}} = x_{\text{bot op}}, \quad y_{\text{top op}} = y_{\text{bot op}} \tag{19.166}$$

Subtracting Equation 19.164 from Equation 19.165 gives:

$$y(\bar{V} - V) = (\bar{L} - L)x - (Dx_D + Bx_B) \tag{19.167}$$

From the overall mass balance around the entire column (see Figure 19.5), where

$$Fx_F = Dx_D + Bx_B \tag{19.168}$$

$$y = -\left(\frac{\bar{L} - L}{V - \bar{V}}\right)x + \frac{Fx_F}{V - \bar{V}} \tag{19.169}$$

Equation 19.169 gives a form of the feed equation since L, $\bar{L}$, V, $\bar{V}$, F, and x_F are constant. It represents a straight line (the feed line) on a McCabe Thiele diagram. The case where the feed flashes in the column to form a vapor and a liquid phase as illustrated in Figure 19.37. Part of the feed, V_F vaporizes while the remainder is liquid, L_F. From Equation 19.169, $(\bar{L} - L)$ is the change in liquid flow rates at the feed stage. That is

$$L_F = \bar{L} - L \tag{19.170}$$

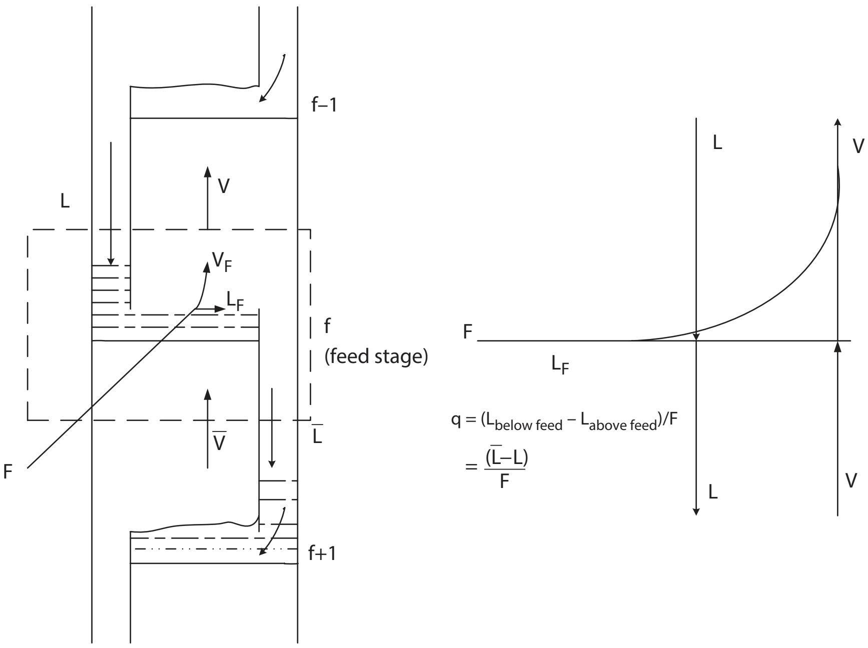


Figure 19.37 Two-phase feed.

The change in vapor flow rate is:

$$V_F = V - \bar{V} \tag{19.171}$$

Equation 19.169 becomes

$$y = -\frac{L_F}{V_F}x + \frac{F}{V_F}x_F \tag{19.172}$$

Equation 19.172 represents the flashing of the feed into the column, and can be written in terms of the fraction vaporized as:

$$y = -\frac{1-f}{f}x + \frac{x_F}{f} \tag{19.173}$$

or in terms of the fraction of remaining liquid, $q = L_F/F$

$$y = \frac{q}{q-1}x + \frac{x_F}{1-q} \tag{19.174}$$

The slope of a line from the intersection point of the feed composition, x_F, with the 45° line on Figure 19.5 is given by $q/(q - 1) = -q/(l - q)$. Physically this gives a good approximation of the number of mols of saturated liquid that will form on the feed plate by the introduction of the feed, although under some thermal conditions the feed may vaporize liquid on the feed plate rather than condense any.

Liebert [80] studied feed preheat conditions and the effects on the energy requirements of a column. This is an essential consideration in the efficient design of a distillation system. As an alternative to locating the "q" line, any value of x_i may be substituted in the "q" line equation below, and a corresponding value of y_i determined, which when plotted will allow the "q" line to be drawn in. This is the line for SV–I, V–I, PV–I, BP–I, and CL–I of Figure 19.36.

$$y_i = -\frac{q}{1-q}x_i + \frac{x_F}{1-q} \tag{19.175}$$

Equation 19.174 or 19.175 is termed the q-line equation. It is located graphically as follows. When $x_i = x_F$, Equation 19.175 gives $y = x_F$. Thus, the q-line passes through the point (x_F, x_F). The slope of the q-line is then calculated or one other convenient point.

A more general definition of q is applied to determine the values of q for subcooled liquid and superheated vapor:

q = enthalpy change to bring the feed to a dew-point vapor temperature divided by enthalpy of vaporization of the feed (dew-point minus bubble-point), i.e.,

$$q = \frac{(H_F)_{\text{Sat'd vapor temperature}} - (h_F)_{\text{feed temperature}}}{(H_F)_{\text{Sat'd vapor temperature}} - (h)_{\text{Sat'd liquid temperature}}} \tag{19.176}$$

where

H_F = feed enthalpy at saturated vapor temperature, kcal/kg, Btu/lb mol.
h_F = feed enthalpy (liquid, vapor or two-phase), kcal/kg, Btu/lb mol.
h = liquid enthalpy, kcal/kg, Btu/lb mol

For a subcooled liquid feed, Equation 19.157 becomes

$$q = \frac{(H_{F,\text{Sat'd vapor temp.}} - h_{\text{Sat'd liquid temp.}}) + (h_{\text{Sat'd liquid temp.}} - h_{F,\text{feed temp.}})}{(H_{F,\text{Sat'd vapor temp.}} - h_{\text{Sat'd liquid temp.}})} \tag{19.177}$$

where

$$(H_{F,\text{Sat'd vapor temp.}} - h_{\text{Sat'd liquid temp.}}) = \Delta H^{vap} \tag{19.178}$$

and

$$(h_{\text{Sat'd liquid temp.}} - h_{F,\text{feed temp.}}) = C_{P_L}(T_b - T_F) \tag{19.179}$$

or

$$q = \frac{\Delta H^{vap} + C_{P_L}(T_b - T_F)}{\Delta H^{vap}} \tag{19.180}$$

For a superheated vapor, Equation 19.176 becomes

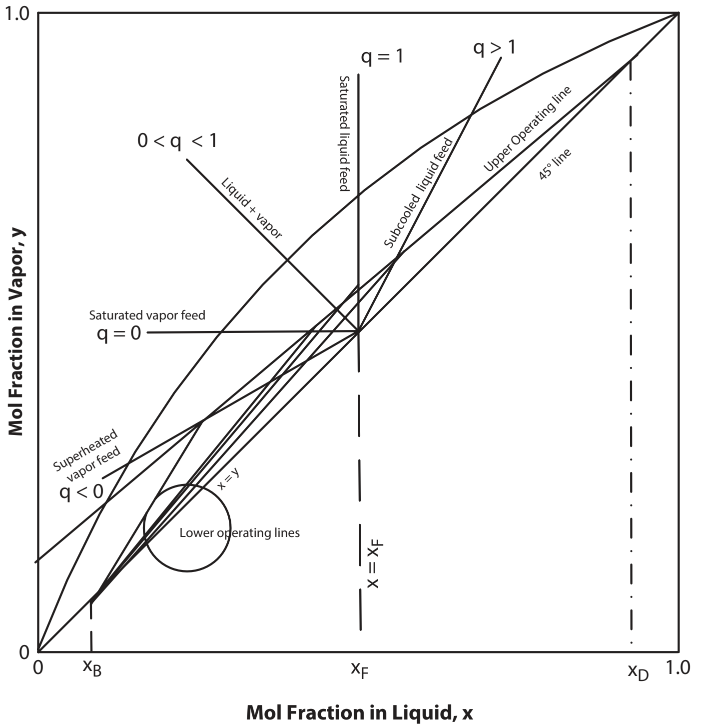


Figure 19.38 Location of q-lines.

$$q = \frac{C_{P_V}(T_d - T_F)}{\Delta H^{vap}} \tag{19.181}$$

where C_{P_L} and C_{P_V} are the liquid and vapor molar heat capacities, respectively, ΔH^{vap} is the molar enthalpy change from the bubble point to the dew point, and T_F, T_d, and T_b are the feed, dew-point and bubble-point temperatures, respectively, of the feed at the column operating pressure. As q changes from a value greater than 1 (subcooled liquid) to a value less than 0 (superheated vapor), the slope of the q-line, q/(q − 1), changes from a positive value to a negative value and back to a positive value. Figure 19.38 shows the effect of thermal condition of the feed on slope of the q-line. For a saturated liquid feed, the q-line is vertical and for saturated vapor, the q-line is horizontal as shown in Figure 19.38.

The relationship between the various types of feed and the slopes of the feed condition is shown in Table 19.22.

19.22 Effect of Total Reflux, Minimum Number of Plates in a Distillation Column

Total reflux exists in a distillation column, whether a binary or multicomponent system, when all the overhead vapor from the top tray or stage is condensed and returned to it. A column may be brought to equilibrium at total reflux for test purposes or for a temporary plant condition, which requires temporary discontinuation of operation. Rather than shut down, drain and then re-establish operating conditions later, it is usually more convenient to maintain a total reflux condition with no feed, no overhead and no bottoms products or withdrawals.

Conditions of total liquid reflux in a column coincide with the minimum number of plates required for a given separation. Under such conditions the column has zero yield of product, and infinite heat requirements, and $L_s/V_s = 1.0$ as shown in Figure 19.36. This is the limiting condition for the number of trays and is a convenient measure of the complexity or difficulty of separation.

Table 19.22 Feed conditions.

Type of feed	**T**	**h_F**	**f**	**q**	**Slope of q-line q/(q-1)**
Subcooled liquid	$T_F < T_{BP}$	$h_F < h$	$f < 0$	$q > 1.0$	> 1.0
Saturated liquid	$T_F = T_{BP}$	h	0	1.0	∞
Liquid-vapor mixture	$T_{DP} > T_F > T_{BP}$	$H_F > h_F > h$	$0 < f < 1$	$1.0 > q > 0$	Negative
Saturated vapor	$T_F = T_{DP}$	H_F	1	0	0
Superheated vapor	$T_F > T_{DP}$	$h_F > H_F$	$f > 1$	$q < 0$	1 > slope > 0

T_{BP} = bubble point of feed; T_{DP} = dew point of feed.

19.22.1 Fenske Equation: Short-Cut Prediction of Overall Minimum Total Trays in a Column With Total Condenser

$$S_m = (N_{min} + 1) = \frac{\log\left(\frac{x_{Dl}}{x_{Dh}}\right)\left(\frac{x_{Bh}}{x_{Bl}}\right)}{\log \alpha_{avg}} \tag{19.182}$$

This includes the bottoms reboiler being equivalent to a tray in the system.

Refer to tabulation below.

N_{min} includes only the required trays in the column itself, and not the reboiler.

$$\alpha_{avg} = (\alpha_{lk/hk})_{avg}$$

D refers to overhead distillate; B refers to bottoms

$$S_m = N_{min} = \frac{\log\left[(x_{lk}/x_{hk})_D(x_{hk}/x_{lk})_B\right]}{\log(\alpha_{lk/hk})_{avg}} \tag{19.183}$$

This applies to any pair of components. Ludwig [81] suggested adding +1 theoretical tray for the reboiler, thus making the total number of theoretical trays a conservative estimate. Converting to actual number of trays requires the calculated tray efficiency:

$$S_m + 1 = N_{min} \tag{19.184}$$

For a condition of overall total trays allowance is to be made for feed tray effect; *add one more theoretical tray to the total.* As demonstrated below, allowance should be made for the reboiler and condenser.

		Total Condenser	**Reboiler**	**Partial Condenser**	**Total**
S	N_{min}	+1	0	+0	$N_m + 1$
	N_{min}	+1	+1	+0	$N_m + 2$
	N_{min}	+0	+1	+1	$N_m + 2$

The approach recommended here is not in agreement with van Winkle [82], who assumes the reboiler and partial condenser are included in the overall calculation for N_{min}.

Various average values of α for use in these calculations are suggested in the following section on "Relative Volatility." Because the feed tray is essentially non-effective, an additional theoretical tray may be added to allow for this. This can be conveniently solved by the nomographs [83] of Figures 19.39 and 19.40. The minimum number of trays in the rectifying section can, if necessary, be calculated by the Fenske equation substituting the limits of x_{Fl} for x_{Bh} and x_{Bl}, and the stripping section can be calculated by difference.

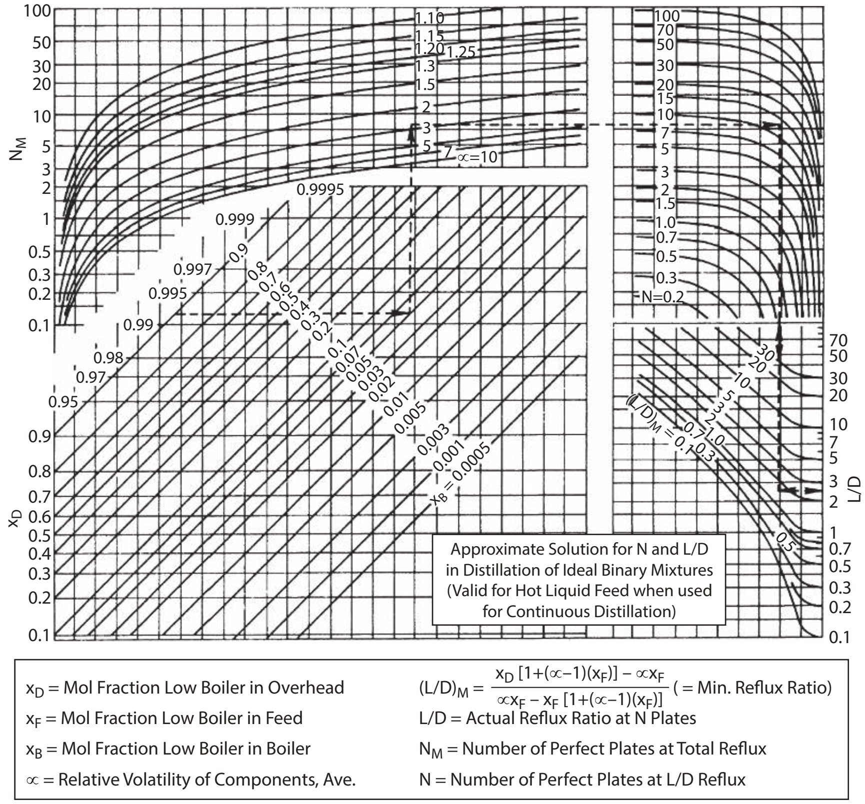


Figure 19.39 Approximate solution for N and L/D in distillation of ideal binary mixtures. Used by permission, Faasen, J. W., Industrial & Eng. Chemistry, V. 36, p. 248 (1944). The American Chemical Society, all rights reserved.

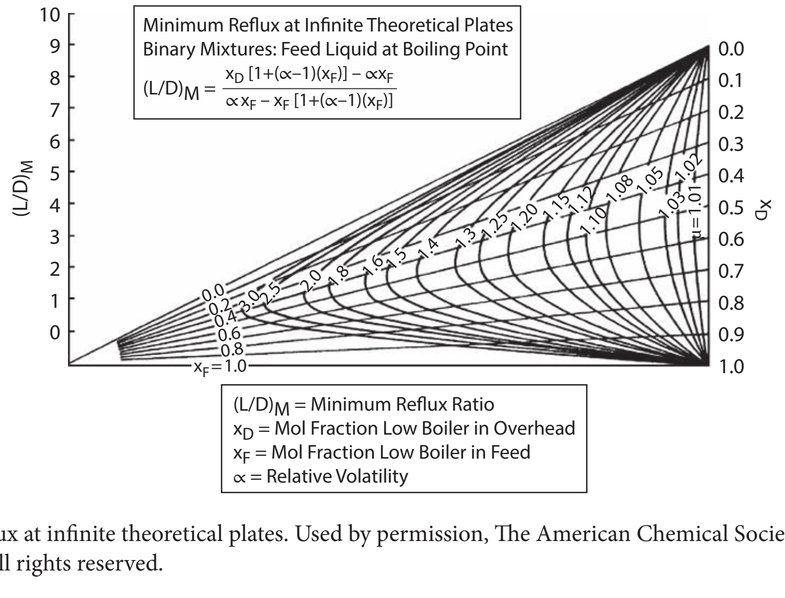


Figure 19.40 Minimum reflux at infinite theoretical plates. Used by permission, The American Chemical Society, Smoker, E. H., Ind. Eng. Chem., V. 34, p510 (1942). All rights reserved.

From Fenske's equation, the minimum number of equilibrium stages at total reflux is related to their bottoms (B) and distillate or overhead (D) compositions using the average relative volatility, see Equation 19.183.

To solve for the component split [84] in distillate or bottoms:

$$\left(\frac{x_{LK}}{x_{HK}}\right)_D = \left(\frac{x_{LK}}{x_{HK}}\right)_D (\alpha_{LK-HK})^{S_m} \tag{19.185}$$

where

S_m = total number of calculated theoretical trays at total reflux, from Equation 19.182
x_{lk} = x_{LK} = liquid mol fraction of light key
X_{hk} = x_{HK} = liquid mol fraction of heavy key
lk - hk = LK - HK = average relative volatility of column (top to bottom)

Clearly, a column cannot operate at total reflux and produce net product. Hence, a reflux ratio of about 1.1 to 1.5 *times the minimum reflux* will usually give practical results. As the reflux ratio is reduced, the number of theoretical trays increases and an economic balance is therefore involved.

19.23 Relative Volatility α Separating Factor in a Vapor–Liquid System

The relative volatility is a measure of the ease of separation, and is the volatility separation factor in a vapor-liquid system, i.e., the volatility of one component divided by the volatility of the other and can be defined as:

$$\alpha_{1-2} = \frac{K - \text{value of component 1}}{K - \text{value of component 2}} \tag{19.186}$$

It represents the ratio between the tendency to vaporize the two components. It also quantifies the tendency for one component in a liquid mixture to separate from the other upon distillation. The term can also be expressed as the ratio of vapor pressure of the more volatile to the less volatile component in the liquid mixture, and therefore $\alpha_{1,2}$ is always equal to ≥ 1.0. If $\alpha_{1,2}$ is high, therefore component 1 has a greater tendency to vaporize (i.e., is more volatile than component 2) and it will be easy to achieve separation by vaporizing one from the other by distillation. However, when component 2 has as high a tendency to vaporize as component 1, the relative volatility will approach unity and the components will be difficult to separate one from the other by distillation. If under the specific conditions $\alpha_{1.2} = 1$, the component cannot be separated by distillation. The greater the value of α the easier will be the desired separation. Relative volatility can be calculated between any two components in a mixture; binary or multicomponent. One of the substances is chosen as the reference to which the other component is compared.

Definition of Relative Volatility: Relative Volatility of Component 1 with respect to component 2.

$$\alpha_{1,2} = (p_1 x_2)/(p_2 x_1) = (y_1 x_2)/(y_2 x_1) \tag{19.187}$$

where

1, 2, etc. are component identification
p = vapor pressure of a component
x = liquid mol fraction of a component
y = vapor mol fraction of a component
π = system total pressure, absolute
Partial pressure:

$$p_a = \pi y_a \tag{19.36}$$

$$p_b = \pi y_b$$

When temperature is constant and at equilibrium for a homogeneous mixture (such as an azeotrope), the composition of the liquid is identical with the composition of the vapor, thus $x_i = y_i$, and the relative volatility is equal to 1.0.

$$K_i = y_i / x_i, \text{that is,} \frac{\text{mol fraction of i in vapor phase}}{\text{mol fraction of i in liquid phase}} \tag{19.38}$$

$$\alpha_{ab} = K_a / K_b = \text{relative volatility of components a to b} \tag{19.39}$$

where

i = compound identification
r = reference compound

As previously discussed, the charts of K values are available, apply primarily to hydrocarbon systems. Reference [79] presents other important data on K value relationships. See Figures 19.11a and 19.11b for charts with pressure effects included. (These are not ideal, but practical charts).

$$\alpha_{1,2} = K_1 / K_2 = P_1 / \pi \tag{19.39a}$$

For multicomponent mixtures [79, 59]:

$$y_2 = \frac{x_2}{\alpha_{1/2} x_1 + x_2 + \alpha_{3/2} x_3 + \alpha_{4/2} x_4 + \ldots} = \frac{x_2}{\sum \alpha x} \tag{19.188}$$

where

1, 2, 3, 4, ...are components in a multicomponent mixture.

$\alpha_{1/2}$ = relative volatility of component 1 with respect to component 2

$\alpha_{3/2}$ = relative volatility of component 3 with respect to component 2.

$$x_2/y_2 = \alpha_{1/2}(x_1/y_1) \tag{19.189}$$

$$y_1 = \alpha_{1/2}x_1\left(\sum \alpha x\right) \tag{19.190}$$

$$y_3 = \alpha_{3/2}x_3\left(\sum \alpha x\right), \text{etc.} \tag{19.191}$$

$$x_1 = \frac{y_1/\alpha_{1/2}}{\sum (y/\alpha)}, x_2 = \frac{y_2/\alpha_{2/2}}{\sum (y/\alpha)}, x_3 = \frac{y_3/\alpha_{3/2}}{\sum (y/\alpha)} \tag{19.192}$$

For a binary system with constant relative volatile:

$$\alpha_{1,2} = \frac{y_1/x_1}{(1-y_1)/(1-x_1)} \tag{19.193}$$

or

$$y_1 = \frac{\alpha x_1}{[1+(\alpha-1)x_1]} \tag{19.194}$$

Winn [72] proposes a modification to allow for temperature variation effects on relative volatility. The method does not apply to mixtures forming azeotropes or at conditions near the critical. Kister [22] proposes:

$$K_1 = \alpha_{1,2}K_2$$

$$N_{min} = \frac{\ln\left[\frac{(x_{lk})_D}{(x_{lk})_B}\left(\frac{(x_{hk_1})_B}{(x_{hk})_D}\right)\right]^{b_{lk}}}{\ln \beta_{lk/hk}} \tag{19.195}$$

α can vary with temperature, so some average α should be used between top and bottom temperature.

When b_{lk} and $\beta_{lk/hk}$ are constants at a fixed or constant pressure, but evaluated for the light (1) and heavy (h) keys at top and bottom temperatures, their relationship is [22]:

$$\beta_{lk/hk} = K_{lk}/(K_{hk})^{b_{lk}}, \text{ at fixed pressure} \tag{19.196}$$

Winn's equation reduced to Fenske's at $b_{lk} = 1.0$ and

$$\beta_{lk/hl} = \alpha_{lk/hk} \tag{19.197}$$

Example 10.14: Determination of Minimum Number of Trays by Winn's Method

The minimum number of trays necessary to debutanizer the effluent from an alkylation reactor will be calculated [72]. The feed, products, and vapor–liquid equilibrium constants of the key components at conditions of temperature and pressure corresponding to the top tray and reboiler are shown in Table 19.23.

The constants b and b are evaluated using Equation 19.196 as follows:

$$0.94 = \beta\,(0.70)^{b}$$

$$3.55 = \beta\,(3.00)^{b}$$

Divide to solve for value of b. Then:

$$3.78 = (4.29)^{b}$$

$$b = 0.913$$

$$\beta = 1.301$$

By use of Winn's Method [72] for product rates:

Table 19.23 Data for alkylation of deisobutanizer; Example 19.14 using Winn's method.

Component	Feed, moles	Over-head, moles	Bottoms, moles	Equilibrium K's	
				Top tray	Reboiler
Ethylene	1	1			
Ethane	2	2			
Propane	48	48			
Isobutene	863	848	15	0.94	3.55
n-Butane	132	71	61	0.70	3.00
Isopentane	33		33		
n-Pentane	5		5		
Alkylate	277		277		
	1361	970	391		

Used by permission, Winn. F.W., *Pet. Ref.* V. 37, No. 5 (1958). P.216. Gulf Pub. Co.

$$\beta^{n+1} = \left(\frac{L_D}{W}\right)\left(\frac{W'}{L'_D}\right)^{b}\left(\frac{B}{D}\right)^{1-b}, \text{ for liquid overhead product} \quad (19.198)$$

$$\beta^{n+2} = \left(\frac{V_D}{W}\right)\left(\frac{W'}{V'_D}\right)^{b}\left(\frac{B}{D}\right)^{1-b}, \text{ for vapor overhead product} \quad (19.199)$$

where

B = mols of bottoms
b = exponent in Equation 19.196
D = total mols of overhead product
n = minimum number of equilibrium trays in tower
$K = y/x$ = vapor–liquid equilibrium ratio for a component
L = mols of a component in liquid phase
P = vapor pressure, psia
T = absolute temperature, °R (=°F + 460)
V = mols of a component in vapor phase
W = total mols of bottoms product
x = mol fraction of a component in liquid phase
y = mol fraction of a component in vapor phase
α = relative volatility
β = constant in Equation 19.195
π = total pressure, psia
L = total mols in liquid phase
V = total mols in vapor phase
subscripts or superscripts:
D = distillate
B = bottoms
(′) = heavy key component
1, 2... = tray number

The minimum number of theoretical stages is calculated as follows:

$$(1.301)^{n+1} = (848/15)(61/71)^{0.913}(391/970)^{0.087}$$

$$(1.301)^{n+1} = 45.5$$

$$n + 1 = 14.5$$

This is exactly the number of stages obtained by tray-to-tray calculations with the K correlation of Winn [74]. The minimum number of stages by the Fenske equation, with a geometric average α of 1.261, is 16.8. Hence the Fenske equation gives an answer greater by 2.3 stages or 16%.

For ideal systems following Raoult's Law; relative volatility $\alpha_{lh} = p_l/p_h$, ratio of partial pressures.

For a binary distillation, α is calculated at top and bottom conditions and a geometric mean used where the differences are relatively small.

$$\alpha_{avg} = \sqrt{\alpha_D(\alpha_B)} \tag{19.200}$$

Kister [22] recommends that the determination of α for calculation is:

(1) α_{avg} = α evaluated at $T_{avg} = (T_{top} + T_{Bot})/2$

$$\text{where } T = {}^\circ F \tag{19.201}$$

(2) α_{avg} = α at feed tray temperature
(3) Winn's [72] method previously discussed

For hydrocarbon systems; the following is often used [49]:

$$\alpha_{ir} = \frac{y_i x_r}{y_r x_i} = \frac{K_i}{K_r} \tag{19.202}$$

where

i = any component
r = component to which all the relative volatilities are referred
K_i = equilibrium distribution coefficient for component, i
K_r = equilibrium distribution coefficient for component to which relative volatilities are referred

For values of α ≈ 1.0, care must be taken in establishing data, as a small change in the value of α_{avg} may double the number of trays.

The exact procedure is to estimate a temperature profile from top to bottom of the column and then calculate α for each theoretical tray or stage by assuming a temperature increment from tray to tray. For many systems this, or some variation, is recommended to perform accurate separation calculations.

$$\alpha_{lh} = \frac{y_l x_h}{x_l y_h} \tag{19.203}$$

For non-ideal systems:

$$\alpha_{lh} = \frac{\gamma_l K_l}{\gamma_h K_h} \tag{19.204}$$

The vapor–liquid equilibrium relationship may be determined from

$$y_l = \frac{\alpha_{lh}(x_l)}{1 + (\alpha_{lh} - 1)x_l} \tag{19.205}$$

By assuming values of x_l, the corresponding y_l may be calculated.

For hydrocarbon systems, where $K_i = y_i/x_i$, then,

$$y_1/y_2 = (K_1/K_2)\,(x_1/x_2),\ \text{and} \tag{19.206}$$

$$\alpha_{i,r} = K_i/K_r = \text{relative volatility} \tag{19.207}$$

Example 19.15: Construction of Boiling Point Curve and Equilibrium Diagram for Benzene–Toluene Mixture

Using the vapor pressure data for benzene and toluene [85]:

1. Construct a boiling point diagram at a total pressure of 760 mm Hg, Figure 19.41.
2. From the boiling point diagram construct the equilibrium x-y curve for a total pressure of 760 mm Hg, Figure 19.42.

Vapor pressure data as read from tables or graphs;

Temperature °C	Vapor Pressure Benzene, mm Hg	Vapor Pressure Toluene, mm Hg
80	760	280
90	1000	410
100	1320	550
110	1740	740
111.5	1760	760

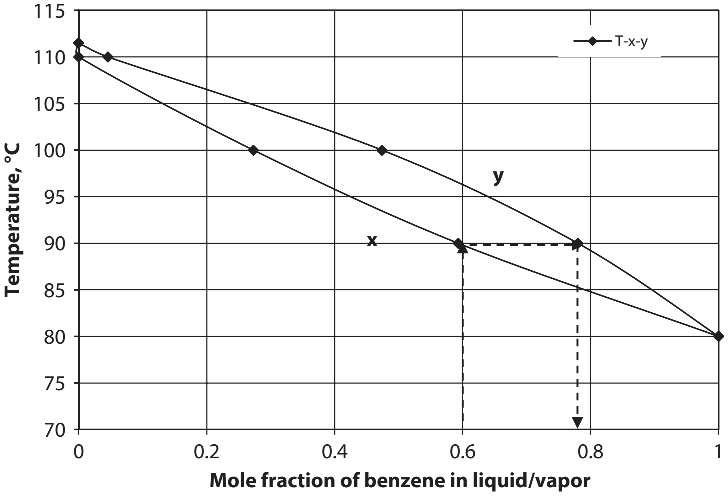


Figure 19.41 Boiling point diagram for Example 19.15. Benzene–toluene mixture, total pressure = 760 mm Hg.

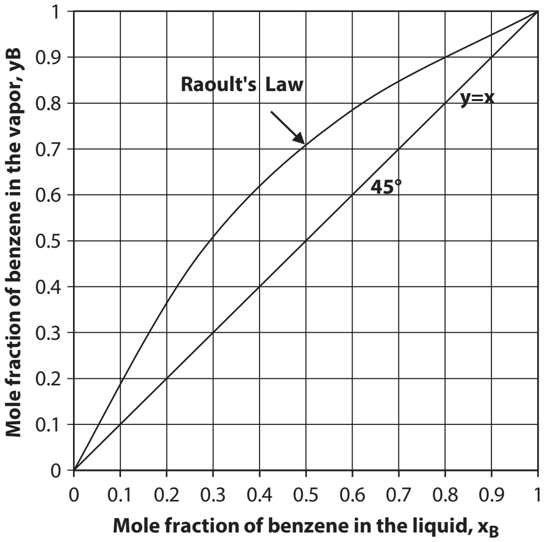


Figure 19.42 VLE diagram for the mixture benzene and toluene at 760 mm Hg total pressure.

Solution

Use Raoult's and Dalton's Laws:

$$p_B = P_B x_B$$

$$p_T = P_T x_T$$

Then: $x_B = \dfrac{\pi - P_T}{(P_B - P_T)}$

$$x_T = 1 - x_B$$

$$y_T = 1 - y_B$$

$$y_B = p_B/\pi$$

where

P_B = vapor pressure, benzene
P_T = vapor pressure, toluene
π = Total pressure

Temp, °C	$(P_B - P_T)$	$(\pi - P_T)$	x_B	x_T	$y_B = \dfrac{P_B x_B}{\pi}$	y_T
80	480	480	1.0	0	1.0	0
90	590	350	0.593	0.407	0.780	0.220
100	770	210	0.273	0.727	0.474	0.526
110	1000	20	0.0	0.980	0.0457	0.954
111.5	1000	0	0.0	1.0	0.0	1.0

Plot values of x_B, x_T, y_T, and y_B on Figure 19.42.

Example 19.16: Repeat Example 19.15 Using K-Values. Refer to Basis of Example 19.14

Using the data from Reference [85] (pp. 221, and 233):

Temp. °C	K_T	**Relative volatility** $\alpha_{B/T}$*	$K_B = \alpha K_T$	$1 - K_T$	$(K_B - K_T)$
80	0.37	2.65	0.981	0.63	0.611
90	0.53	2.48	1.314	0.47	0.784
100	0.73	2.39	1.745	0.27	1.015
110	0.97	2.35	2.28	0.03	1.31
111.5	1.0	2.35	2.35	0.0	1.35

*Read from chart [85].

$X_B = \dfrac{(1-K_T)}{(K_B-K_T)}$	$x_T = 1 - x_B$	$y_B = K_B x_B$	$y_T = K_T x_T$
*1.031	0.0	*0.981	0.0
0.60	0.40	0.789	0.212
0.266	0.734	0.464	0.535
0.0229	0.977	0.0523	0.946
0.0	1.0	0.0	1.0

*Note: If graphs could be read close, these values would be equal to 1.0.

Procedure

1. Read K value for toluene from tables or charts.
2. Read α values for benzene/toluene from Reference [85].
3. Calculate K (benzene) from: $\alpha_{B/T} = K_B/K_T$
4. Calculate $x_{benzene}$:

$y_B = K_B x_B$
$y_T = K_T x_T$
$\Sigma 1.0 = K_B X_B + K_T x_T = K_B x_B + K_T (1 - X_B)$ $= X_B (K_B - K_T) + K_T$
$X_B = (1 - K_T)/(K_B - K_T)$

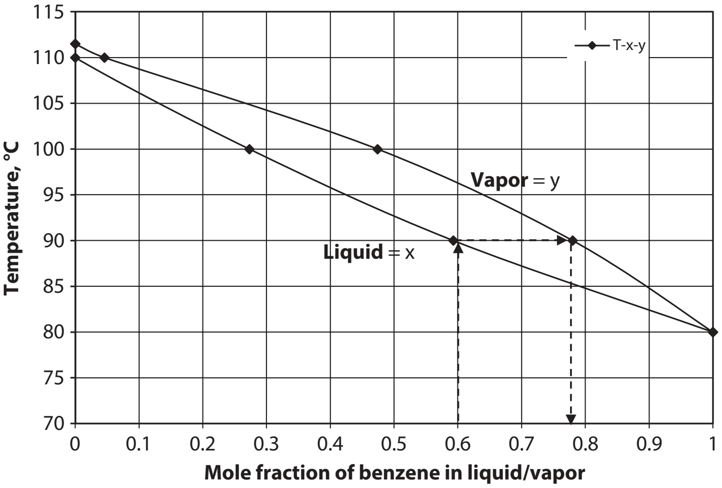


Figure 19.43 Boiling point diagram for benzene–toluene mixture using K values, total pressure 760 mm Hg for Example 19.16.

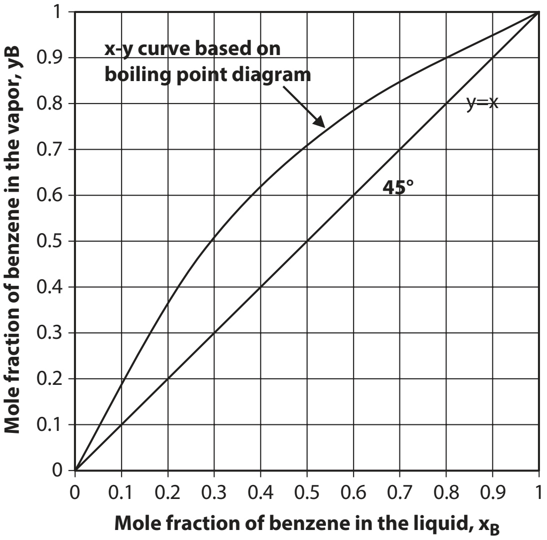


Figure 19.44 x-y diagram for benzene–toluene mixture, 760 mm Hg pressure, based on K-values, Example 19.16.

5. Calculate $y_{benzene}$: $y_B = K_B x_B$
6. Plot boiling point diagram, see Figure 19.43.
7. Plot x-y diagram, see Figure 19.44.

Example 19.17: Calculation of Flash Vaporization of a Hydrocarbon Liquid Mixture

What fraction of a liquid mixture containing 10 mole % propane, 65% n-butane, and 25% n-pentane would be vaporized in a flash vaporization process at a temperature of 40°F and a pressure of 600 mm Hg abs?

The following vapor pressure data for 40°F are available:

propane	3800 mm Hg
n-butane	820 mm Hg
n-pentane	190 mm Hg

Solution

Assume Raoult's Law is applicable. At a total pressure, π, the temperature of flash must be between the dew point and bubble point.

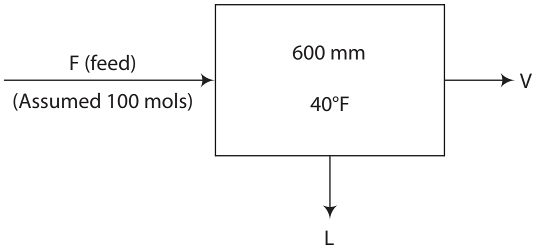
Raoult's Law:

$p_1 = P_1 x_1$

$K_1 = y_1/x_i$

$py_1 = P_1 x_1$

$y_1/x_1 = P_1/p = K_1$



	Mol%	**Vapor Press. Mm Hg = P_1**	$K = \frac{P_1}{600}$	**FX**	$\frac{L}{V}$	$\frac{L}{KV}$	$1 + \frac{L}{KV}$
C_3H_8	10	3800	6.34	10		0.158	1.158
nC_4H_{10}	65	820	1.37	65	1.0	0.73	1.73
nC_5H_{12}	25	190	0.317	25		3.16	4.16
$V = \frac{FX_i}{1 + L/(K_i/V)}$							
8.64							
37.6							
+ 6.0							

V = 52.24 NOT a check, reassume and recalculate.

See Figure 19.45 for plot of results and the resultant extrapolation. This type chart serves as a guide to reduce the number of "guesses" to reach an acceptable solution. After several assumptions:

1. Assume feed = 100 mols
2. Assume L = 30
3. Then: V= 100 – 30 = 70 mols

X = mols of component i in vapor plus mols of component i in liquid divided by the total mols of feed (liquid + vapor)

F = mols of feed

Following the same headings as previous table it continues:

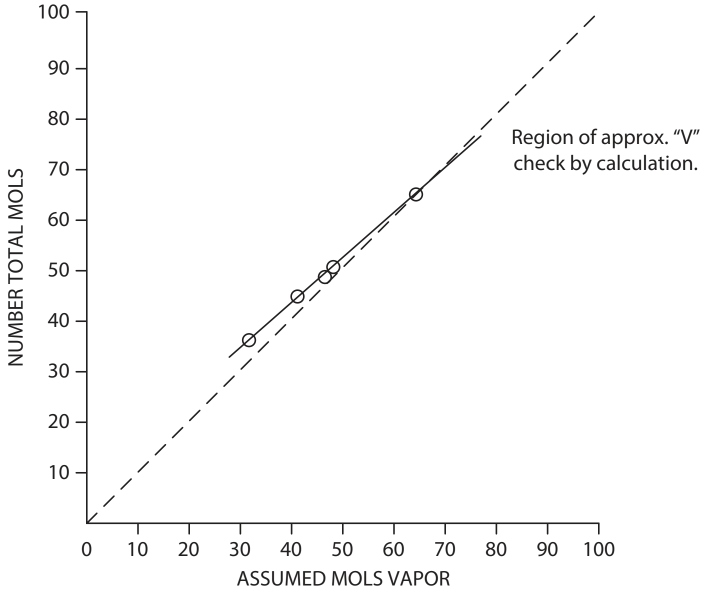


Figure 19.45 Extrapolation curve to determine approximate value of "V" for Example 19.17.

	Mol%	FX	Vapor Press.	$\frac{L}{V}$	$\frac{L}{KV}$	$1+\frac{L}{KV}$
C_3H_8	10	10	3800	0.429	0.0677	1.067
n-C_4H_{10}	65	65	820	0.429	0.314	1.314
n-C_5H_{12}	25	25	190	0.429	1.352	2.352
$V=\frac{FX}{1+\frac{L}{KV}}$		*y; mol. Frac.				
9.38		0.135				
49.4		0.711				
10.63		0.153				
69.41		0.999				

These values are sufficiently close enough for most calculations. Therefore, after several trial-and error calculations these results indicate that after flashing, there would be approximately 70% vapor of above composition and 30% (mol) liquid.

19.24 Rapid Estimation of Relative Volatility

Wagle [86] presents an estimate method for the average relative volatility of two components, related to their normal boiling points and the latent heats of vaporization within the temperature range of their individual boiling points:

$$\alpha = \exp\,[0.25164\,(1/T_{bl} - l/T_{b2})\,(L_1 + L_2)] \qquad (19.208)$$

where

α = relative volatility between the two components in the temperature range T_{bl} to T_{b2}
T_{bl} = normal boiling point of Component l, K
T_{b2} = normal boiling point of Component 2, K
L_l = latent heat of vaporization for Component 1 at T_{bl} kcal/kmole
L_2 = latent heat of vaporization for Component 2 at T_{b2}, kcal/kmole

If a compound's latent heat is not known, it can be estimated from the normal boiling point and molecular weight.

Example 19.18: Estimation of Relative Volatility by Wagle's Method

The average relative volatility of benzene and toluene can be determined using the following data: T_{bb} = 353.3 K, T_{bt} = 383.8 K, L_b = 7352 kcal/kmole, and L_t = 7930 kcal/kmole (where the subscripts b and t denote benzene and toluene, respectively) [86]. Substituting these values into Equation 19.208 above gives:

$$\alpha_{bt} = \exp\left[0.25164\times\left(\frac{1}{353.3}-\frac{1}{383.8}\right)\times(7{,}352+7{,}930)\right] = 2.375$$

This compares with a value of 2.421 for α determined using vapor-pressure/temperature charts.

19.25 Estimation of Relative Volatilities Under 1.25 (α < 125) by Ryan

For key components with relative volatilities < 1.25, McCabe–Thiele diagram is difficult to create because the stage equilibrium data and the operating line data are very close to the x-y diagonal line, and the graph is indecipherable [14]. Ryan [14] recently developed an alternative diagram in which the transformation spreads out the region close to the diagonal over the whole height of the plot. This approach provides more insight into the separation of iso-butane and n-butane as shown in Table 19.24.

The procedures for creating an alternative diagram are as follows:

1. Plot (y – x) versus x instead of y on the y-axis.
2. The operating lines begin on the x-axis at x_B and x_D. Both slopes are [(L/V) – 1]. The rectifying line slope will be negative and the stripping line slope positive.

Table 19.24 Components with relative volatilities less than 1.25 (α < 1.25).

Components	Relative volatility, α
Butane splitters in natural-gas–liquid plants and refineries	about 1.25
De-isobutanizers in butylenes alkylation units in refineries	about 1.25
Argon columns in cryogenic air plants	1.1 < α < 1.6
Columns to separate ortho, para- and metaxalone in petrochemical plants	about 1.15
Propane-propylene splitters	about 1.1
Ethylbenzene-xylene superfractionators	about 1.05

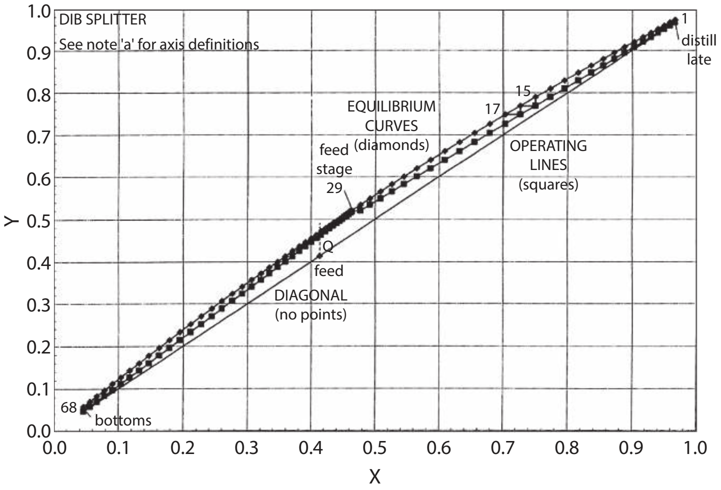


Figure 19.46 McCabe–Thiele diagram.

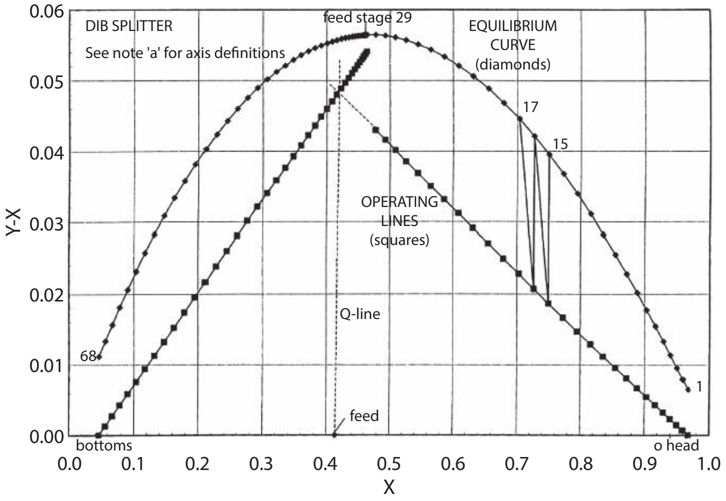


Figure 19.47 Revised McCabe–Thiele diagram.

3. The q-line starts on the x-axis at x_F. The value of q is the same as for conventional McCabe–Thiele. The slope of the q-line in the graph is the McCabe–Thiele slope minus 1. Therefore, for a feed at its bubble point, the q-line is vertical (slope = infinity) for the conventional McCabe–Thiele and Ryan calculations. For the dew point feed, the slope is 0 for the conventional McCabe–Thiele and −1 for Ryan.
4. The familiar steps for the determination of theoretical stages on the McCabe–Thiele diagram are modified on the Ryan plot. The vertical portions remain vertical, but the horizontal portions become slanted with a slope of −1. Figures 19.46 and 19.47 show McCabe–Thiele and Ryan's plots, respectively.

19.26 Estimation of Minimum Reflux Ratio: Infinite Plates

As the reflux ratio is decreased from infinity for the total reflux condition, more theoretical steps or trays are required to complete a given separation. When the limiting condition of Figure 19.48 is reached; where the operating line touches the equilibrium line and the number of steps from the rectifying to stripping sections becomes infinite.

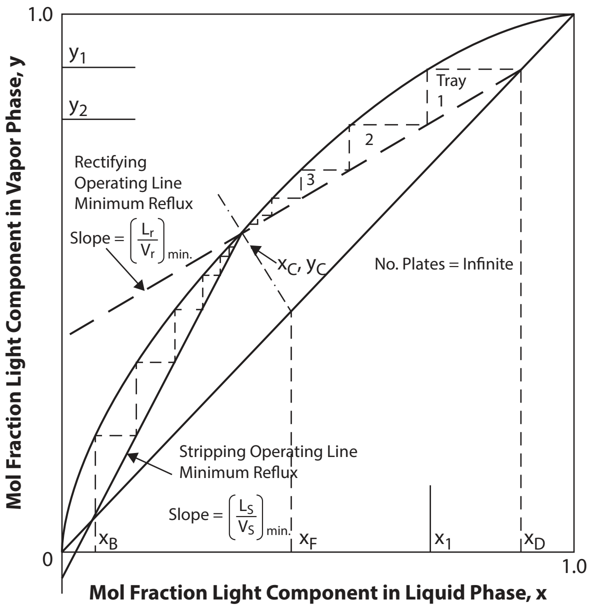


Figure 19.48 Fractionation of binary mixture at minimum reflux condition.

If the operating lines of Figure 19.48 intersect at x_c, y_c outside or above the equilibrium line when insufficient reflux is used, the separation is impossible.

This graphical representation is easier to use for non-ideal systems than the calculation method. This is another limiting condition for column operation, i.e., below this ratio the specified separation cannot be achieved even with infinite number of plates. This minimum reflux ratio can be determined graphically from Figure 19.48, as the line with smallest slope from x_D intersecting the equilibrium line at the same point as the “q” line for mixture following Raoult’s Law.

External reflux ratio = L/D

Slope of line from x_D:

$$\left(\frac{L}{V}\right)_{min} = \frac{(L/D)_{min}}{(L/D)_{min}+1} = \frac{x_D - y_c}{x_D - x_c} \tag{19.209}$$

L/V = internal reflux ratio

For non-ideal mixtures the minimum L/V may be as indicated in Figure 19.36 (abnormal equilibrium), and hence not fixed as indicated above.

Figure 19.40 presents a convenient and acceptably accurate nomogram of Smoker’s [87].

$$\left(\frac{L}{D}\right)_{min} = \frac{x_D - y_c}{y_c - x_c} \tag{19.210}$$

where x_c and y_c are coordinates of intersection of minimum reflux "operating" line with the equilibrium curve. At boiling point $x_c = x_F$.

Underwood's algebraic evaluation [85] for minimum reflux ratio is applicable for handling ideal, or near ideal systems:

For a liquid at its bubble point, q = 1.0

$$\left(\frac{L}{D}\right)_{min} = \frac{1}{\alpha - 1}\left[\frac{x_{lD}}{x_{lF}} - \frac{\alpha(1 - x_{lD})}{(1 - x_{l}F)}\right] \tag{19.211}$$

For high-purity separations where $x_D \sim 1.0$, with reasonably well-balanced columns (where $0.2 < x_F < 0.8$) and moderate values of α,

$$\left(\frac{L}{D}\right)_{min} \approx \frac{1}{(\alpha - 1)x_F} \tag{19.211a}$$

For a vapor feed with no superheat, q = 0

$$\left(\frac{L}{D}\right)_{min} = \frac{1}{\alpha - 1}\left[\frac{\alpha x_{lD}}{x_{lF}} - \frac{(1 - x_{lD})}{(1 - x_{l}F)}\right] - 1 \tag{19.212}$$

For the general case, the relation is more complex in order to solve for $(L/D)_{min}$

$$\frac{(L/D)_{min}(x_{lF}) + qx_{lD}}{(L/D)_{min}(1 - x_{lF}) + q(1 - x_{lD})} = \frac{\alpha[((L/D)_{min} + 1)y_{lF} + (q - 1)x_{lD}]}{[(L/D)_{min} + 1](1 - x_{lF}) + (q - 1)(1 - x_{lD})} \tag{19.213}$$

Short *et al.* [89] discuss minimum reflux for complex fractionators.

19.27 Calculation of Number of Theoretical Trays at Actual Reflux

The Gilliland correlation [90] of Figure 19.49a has proven satisfactory for many binary as well as multicomponent mixtures over a wide range of reflux ratios and number of theoretical plates. Many systems appear to be economically designed for $\frac{(L/D) - (L/D)_{min}}{(L/D) + 1} = 0.1 \text{ to } 0.33$ and with actual reflux ratios of 1.2 to 1.5 times at minimum. This is inapplicable to systems of greatly varying relative volatility. A Ponchon–Savarit or enthalpy method must then be followed.

Eduljee [91] suggests an easier equation to replace the Gilliland plot. The data input is the same. For tray towers:

$$Y_T = 0.75 - 0.75X^{0.5668} \tag{19.214}$$

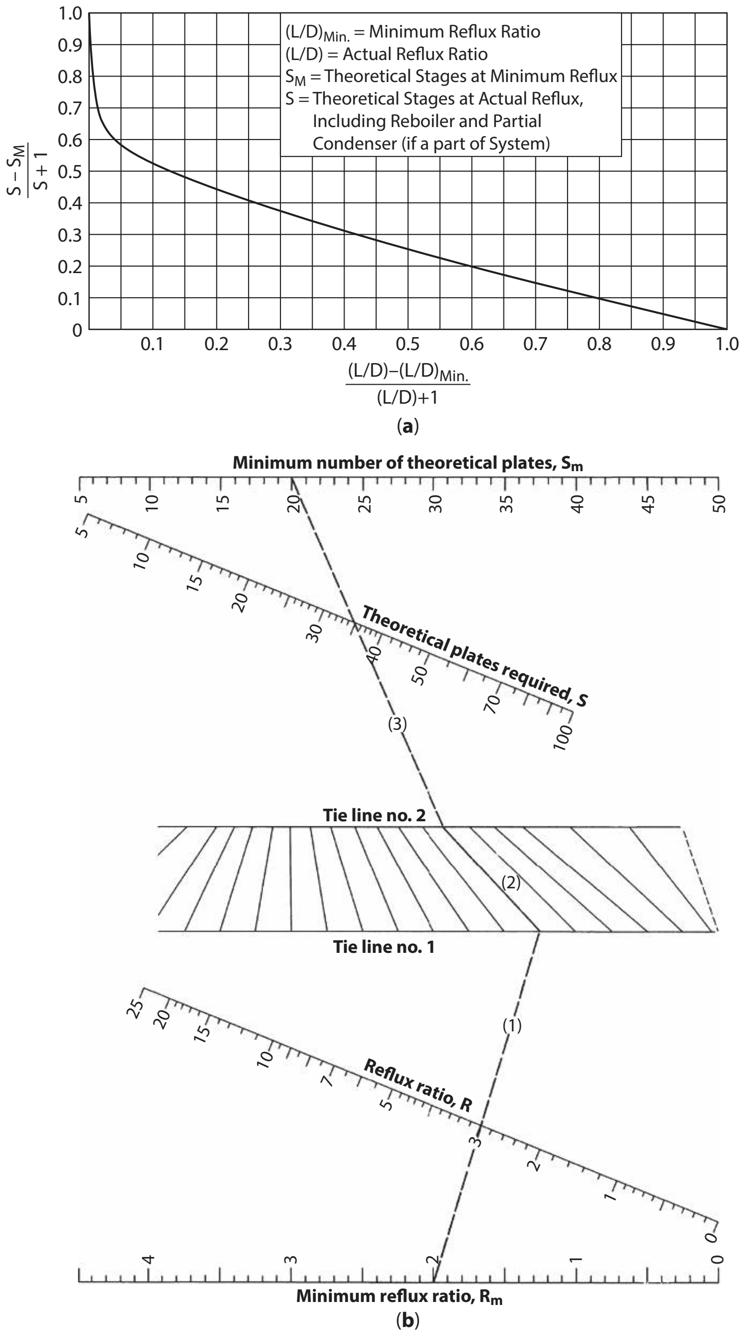


Figure 19.49 (a) Correlation of theoretical plates with reflux ratio, (b) Chart for reflux vs. trays. Use this nomograph for Gilliland's calculations for number of theoretical plates/trays. Used by permission, Mapstone, G. E., Hydrocarbon Processing, V. 47, No. 5, p. 169, (1968). Gulf Publishing Co., All rights reserved.

$$Y_T = \frac{S_t - S_{min}}{(S+1)} \tag{19.215}$$

$$X = \frac{(L/D) - (L/D)_{min}}{(L/D)+1} \tag{19.216}$$

where

S_t = theoretical actual trays at actual reflux, L/D, including overhead total condenser and reboiler
Y_T = correlation expression similar to Gilliland's
X = correlation expression similar to Gilliland's
R = reflux ratio, L/D where L is liquid returned to the column in mols/h
D = distillate rate in mols/h
L = liquid returned to column, mols/h
subscripts:
h = heavy
min = minimum
P = for packed towers
T = for tray column

After calculating X, and solving for Y using Equation 19.214, then solve for the theoretical trays, S_t, at the actual selected reflux ratio (L/D) from the equation for Y. The equation appears to represent several reliable data references. A review of alternative equations to replace the Gilliland plot is provided later in this book.

For packed towers, the corresponding relation for trays is [91]:

$$Y_p = 0.763 - 0.763X^{0.5806} \tag{19.217}$$

and,

$$Y_P = \frac{NTU - NTU_{min}}{NTU + 2\alpha} \tag{19.218}$$

where

NTU = total number of transfer units
$\alpha = \frac{\ln\alpha}{(\alpha - 1)}$, where α_h taken as 1.0

Mapstone [92] and Zankers [93] developed the chart shown in Figure 19.49b to follow Figure 19.49a to allow for a rapid evaluation of Gilliland's equation for the number of theoretical plates at any reflux and minimum number of theoretical plates and minimum reflux ratio. The accuracy appears generally satisfactory for initial industrial design. It can therefore be a useful time-saving device when evaluating a range of values. For another interesting attempt to improve the Gilliland plot by use of equations, refer to Reference [94].

Example 19.19: Using Figure 19.49b to Solve Gilliland's Equation to Determine the Minimum Theoretical Plates for Setting Actual Relux

If the minimum reflux ratio is 2.0 and the minimum number of theoretical plates is 20, how many theoretical plates will be required if a reflux ratio 1.5 times the minimum is used [92]?

Solution

The required reflux ratio, R = 1.5 × 2.0 = 3.0

1. Connect 2.0 on left hand R_{min} scale with 3.0 on left diagonal R scale and extend to cut Tie Line 1.
2. Transfer this value across the central maze to Tie Line 2.
3. Connect this point on Tie Line 2 with 20 on the right hand S_{min} scale to cut the right diagonal S scale at 35 (calc. 34.9).

The number of theoretical plates required will be 35. If any three of the four variables, S, S_{min}, R, and R_{min} are known, this chart can be used by an analogous procedure to give the fourth.

where

S = theoretical plates at any reflux
S_{min} = minimum number of theoretical plates
R = any reflux ratio
R_{min} = minimum reflux ratio

19.28 Identification of "Pinch Conditions" on an x-y Diagram at High Pressure

A column is referred to be "pinching" when the component balance line is too close to the equilibrium curve. This physically represents a situation where several stages in the column are accomplishing very little separation and are thus redundant. This phenomenon not only wastes stages, but could result in an oversized column. Additionally, inaccuracies in relative volatilities and enthalpies may bring the component balance line and the equilibrium curve closer or even cross the equilibrium curve earlier. This could increase the stage requirement, and if the additional stages required are unavailable, the column will not achieve its desired separation.

Wichterle *et al.* [95] identify that near the critical pressure point of the more volatile component, all systems exhibit a "pinch" phenomenon at high pressure as shown in Figure 19.50 [95]. To obtain separation performance, the K-value data used must be accurate in this narrow range of separation. For hydrocarbon systems, as well as systems involving hydrogen, nitrogen, and methane, the data must be accurate and use of a general equation for the particular compound is inadequate. This is crucial for high accuracy in computer performance analysis.

In an existing column, where the feed location is already in position, a pinch can be remedied by increasing the reflux and reboil. This invariably draws the component balance line and the equilibrium curve further apart, enlarging the steps and thereby permitting the required separation to be achieved with the existing feed location. The drawback is lower feed capacity, higher energy consumption, greater vapor and liquid flows through the column, and eventual higher operating costs. Therefore, it is essential to design a column away from the pinched region.

McCormick [96] presents a correlation for Gilliland's chart relating reflux ratio, minimum reflux ratio, number of theoretical stages, and minimum theoretical stages for multicomponent distillation. Selection of a multiplier for actual reflux ratio over minimum reflux ratio is important. Depending on the complexity and analysis of the

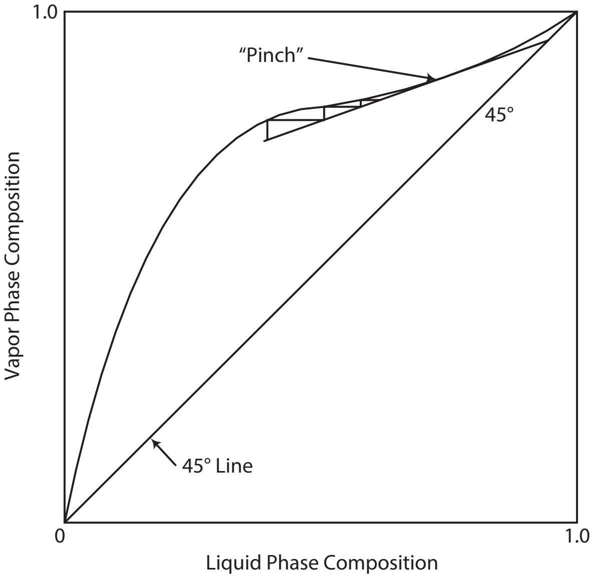


Figure 19.50 Example of typical "pinch point" for critical region for high-pressure distillation. Used by permission, Wichterle, I., Kobayashi, R., and Chappelear, P.S., Hydrocarbon Processing, p. 233, Nov., 1971, Gulf Publishing Co., All rights reserved.

component's separation by the stages, the actual reflux may vary from 1.2 to 1.5 to 2.0, dependent on the economics. The proposed equation agrees satisfactorily with other methods, and especially in the extreme ranges of Gilliland's plots [96], as well as the most used region.

$$R = AR_{min} \tag{19.219}$$

$$X = \frac{R - R_{min}}{R + 1} = \frac{(A-1)(R_{min})}{(AR_{min}+1)} = \frac{A-1}{\left(A + 1/R_{min}\right)} \tag{19.220}$$

Representative values of X calculated from Equation 19.220 are given in the following Table 19.25 for values of R_{min} and multiplier A. Actual reflux ratio values can be assumed, and the system tested for $R_{min.}$

For actual versus minimum number of stages in a column,

$$Y = (N - N_{min})/(N + 1) \tag{19.221}$$

$$Y = 1 + \left[\frac{(R - R_{min})}{(R+1)}\right]^{(0.0456 \ln X + 0.44)} \tag{19.222}$$

$$X = (R - R_{min})/(R + 1) \tag{19.223}$$

where

A = parameter in correlating equation or multiplier on R_{min}
B = parameter in correlating equation
ln = natural logarithm
log = logarithm to the base 10

Table 19.25 Operational values of X calculated via equation 19.220 for a range of reflux ratios*.

R_{min}	**Multiplier A**										
	0.02	**1.05**	**1.07**	**1.10**	**1.12**	**1.15**	**1.17**	**1.20**	**1.30**	**1.40**	**1.50**
1	0.010	0.024	0.034	0.048	0.063	0.070	0.092	0.091	0.130	0.167	0.200
3	0.015	0.036	0.050	0.070	0.092	0.101	0.132	0.130	0.184	0.231	0.273
5	0.016	0.040	0.055	0.077	0.102	0.111	0.145	0.143	0.200	0.250	0.294
10	0.018	0.043	0.060	0.083	0.110	0.120	0.157	0.154	0.214	0.267	0.313
Total reflux	0.020	0.048	0.065	0.091	0.120	0.130	0.170	0.167	0.231	0.286	0.393

*Used by permission, McCormick, J. E., Chemical Engineering V. 95, No. 13 (1988), all rights reserved.

N = actual number of theoretical stages required for a given separation
N_{min} = minimum number of theoretical stages required for a given separation
R = external reflux ratio for a given separation
R_{min} = minimum external reflux ratio for a given separation

$$X = (R - R_{min})/(R + 1)$$

$$Y = (N - N_{min})/(N + 1)$$

The following is a short approximation method for minimum reflux ratios for multicomponent mixtures [97]:

$$R_{min} = \frac{1}{[(x_{FLK})_{eff}][(\alpha_{LK})_{avg} - 1]} \tag{19.224}$$

where

$$(x_{FLK})_{eff} = x_{FLK}/(x_{FLK} + x_{FHK})$$

n = number of components
R_{min} = minimum reflux ratio
x = liquid mol fraction
α_i = relative volatility of component i based on heavy key
α_{LK} = relative volatility of component, i, based on light key
subscripts:
avg = average
eff = effective
F = feed
FHK = heavy key in feed
FLK = light key in feed
i = component
LK = light key
HK = heavy key

Kister [22, 98] examines binary distillation systems with multiple feeds, one or more side products, one or more points of heat removal or addition, and various combinations.

19.29 Distillation Column Design

In designing a distillation column, the chemical/process engineer often is confronted with three basic tasks:

- Choice between tray or packed column
- Design of the column, and its ancillaries, e.g., feed preheater, reflux condenser, reflux splitter, reboiler, and so on.
- Rating problem

In the design of a plate column, it is a basic requirement to calculate how many plates/trays are required to achieve the desired degree of separation of the components from the specified feed conditions. This aspect relates to size and configuration of the equipment. In the rating, the performance of a column under various conditions with a fixed number of plates/trays after it is specified should be predicted.

19.29.1 Design Method for a Plate Column

The following is a design procedure using graphical solution.

Given:

Feed Conditions

- **(i)** Feed flow rate, F
- **(ii)** Feed composition, z_F
- **(iii)** Feed temperature or enthalpy or thermal condition, q
- **(iv)** Column pressure
- **(v)** Reflux temperature or enthalpy (usually saturated liquid)

Product Conditions

$$\left.\begin{array}{l}\text{(i) Distillate composition } x_D\\ \text{(ii) Bottoms composition } x_B\end{array}\right\}\text{ all with respect to light component}$$

Reflux Ratio, R: Relative volatility, α, The vapor-liquid equilibrium curve

Procedure

1. Plot the vapor–liquid equilibrium curve from the data available at the column pressure. In terms of relative volatility:

$$y = \frac{\alpha x}{[1+(\alpha-1)x]} \tag{19.48}$$

where α is the geometric average relative volatility of the lighter (more volatile) component with respect to the heavier component (less volatile).

2. Draw the 45°-line, that is, x = y line.

3. Make a material balance over the column to determine the top and bottom compositions, x_B, x_D from the data given.
4. Locate the points x_B, x_D and z_F on the 45°-line. That is, draw perpendiculars on x-axis at points x_B, x_D and z_F. Their points of intersection with the 45°-line are the required locations.
5. Calculate the distillate and bottoms flow rates.

Distillate Flow Rate:

$$D = F\left[\frac{(z_F - x_B)}{(x_D - x_B)}\right] \tag{19.225}$$

Bottoms flow rate:

$$B = F - D \tag{19.226}$$

6. The point of intersection of the two operating lines depends on the phase condition of the feed. The line on which the intersection occurs is referred to as the q line. The q-line is found as follows:

   (i) Calculate the value of the ratio q given by

$$q = \frac{\text{heat to vaporize 1 mol of feed}}{\text{molar latent heat of feed}} \tag{19.227}$$

where the numerator in Equation 19.227 is the heat to make saturated vapor at the feed tray pressure from 1 mol of feed at its thermal condition, as feed.

   (ii) Plot the q-line, slope = $q/(q-1)$, intersecting the diagonal at z_F (the feed composition).

7. Select the reflux ratio and determine the point where the top operating line extended cuts the y axis:

$$\phi = \frac{x_D}{1+R} \tag{19.228}$$

8. Draw in the top operating line, from x_D on the diagonal ϕ to the y-axis at the value of ϕ.
9. Draw in the bottom operating line; from x_B on the diagonal to the point of intersection of the top operating line and the q-line.
10. Starting at x_D or x_B, step off the number of stages.
11. The design is complete.

Count the number of trays in the rectifying section.

Count the number of trays in the stripping section.

Sum the number of trays in both rectifying and stripping sections.

Note: The feed point should be located on the stage closest to the intersection of the operating lines.

The reboiler and a partial condenser (if used) act as equilibrium stages. Figure 19.5 shows the McCabe–Thiele diagram, and the method can be used for the design of columns with side streams and multiple feeds. The liquid and vapor flows in the sections between the feed and the take-off points are calculated and operating lines drawn for each section.

19.29.2 Continuous Fractionating Column

Rating Method

On completion of the column design, it should be checked to see whether it will perform under different conditions. This rating is an operating problem. Typically, the heat input rate to achieve a specified separation when the feed condition varies will be determined. Another type of rating problem can involve a trial-error iterative solution as follows:

(i) An assumption is made of one or more unknown variables.

(ii) On the basis of these assumed values, the design procedure is carried out.

(iii) The results are compared with the column specifications.

(iv) If they agree, no further assumption is required, and the estimated values of the unknown variables are acceptable, otherwise another set of values are assumed and the design procedure is repeated.

An essential difference between the design and rating methods is that the feed is not introduced on the optimum feed tray, but is obtained through the design calculation. In general, the actual feed tray is not the optimum feed tray. Iterative procedures are essential to apply rating methods to the design of a new column. An initial estimate of the number of stages can be made using short-cut methods and the programs used to calculate the product compositions. Calculations are repeated with revised estimates until a satisfactory design is achieved.

19.30 Simulation of a Fractionating Column

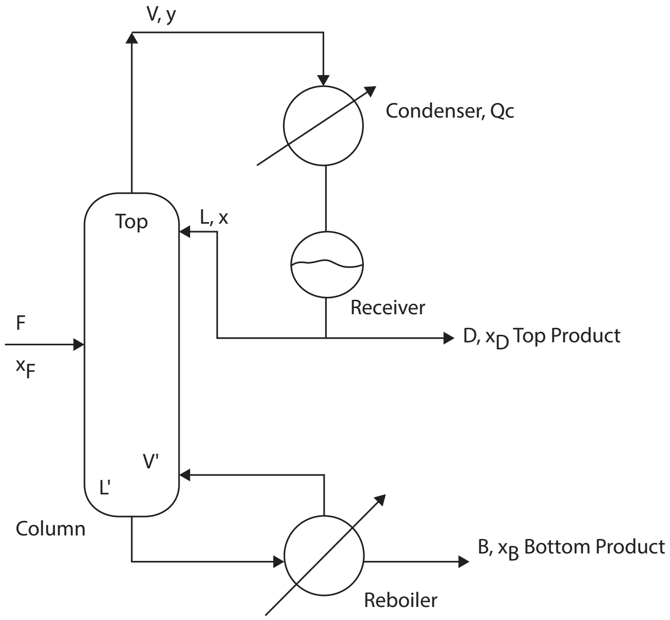
In a simulation exercise, the objective is to predict how much separation can be achieved with a given feed in a column, which has been commissioned or is in use. The number of stages and feed stage location are already fixed, as are the column diameter and the reboiler size; these fix the maximum vapor flow rate. There are various ways of specifying the remaining parameters. The desired composition of more volatile component in the distillate and bottoms could be set; the requirement is to determine the external reflux ratio that will produce the separation. A check may also be made that the maximum vapor flow rate will not be exceeded. Alternatively, the distillate or bottoms compositions and reflux ratio are specified, and then the unknown composition and the vapor flow rate are determined. Another alternative is to specify the heat load in the reboiler and the distillate or bottoms compositions thus requiring the reflux ratio and unknown product compositions to be determined and the vapor flow rate checked. In general, all variables that can be controlled are specified, and these are illustrated by Wankat [99].

Example 19.20

An equimolar binary mixture with a relative volatility of 2 is fed to a distillation column which produces a distillate containing 97.5% of the more volatile component and a residue containing 10% of the more volatile component. The feed is liquid at its boiling point and 80% of the liquid entering the reboiler is vaporized and returned to the column as boil-up.

(i) Estimate the number of theoretical plates provided by this column.
(ii) Check your answer in part (i) by reference to the graph of the Gilliland correlation.

Solution



The overall mass balance on the column.

$$F = D + B \tag{19.229}$$

The component balance on the more volatile component (MVC) is:

$$Fx_F = Dx_D + Bx_B \tag{19.230}$$

Assuming that F = 100 mols/h

$$x_F = 0.5, x_D = 0.975 \text{ and } x_B = 0.1$$

Substituting the values of the components in Equation 10.230 gives:

$$100(0.5) = D\,(0.975) + B\,(0.1) \tag{19.231}$$

$$50 = 0.975D + 0.1B$$

Rearranging Equation 19.229 and substituting into Equation 19.231 gives:

$$50 = 0.975D + 0.1\,(100 - D) \tag{19.232}$$

$$50 = 0.975D + 10 - 0.1D \tag{19.233}$$

$$D = 45.71 \text{ mols/h}$$

$$B = 54.29 \text{ mols/h}$$

The overall mass balance on UOL is:

$$V = L + D \tag{19.234}$$

Component balance is:

$$V_n y_n = L_{n+1} x_{n+1} + Dx_D \tag{19.235}$$

or

$$y_n = \frac{L_{n+1}}{V_n} x_{n+1} + \frac{D}{V_n} x_D \tag{19.236}$$

$$\text{Reflux ratio, } R = \frac{L}{D} \tag{19.237}$$

Substituting Equation 19.237 into Equation 19.234 gives:

$$V_n = RD + D \tag{19.238}$$

Equation 19.236, Upper Operating Line (UOL), assuming constant molar overflow, becomes:

$$y_n = \frac{R}{(R+1)} x_{n+1} + \frac{x_D}{R+1} \tag{19.239}$$

The intercept on the UOL is:

$$I = \frac{x_D}{(R+1)} \tag{19.240}$$

Since the feed is liquid at its boiling point (i.e., q = 1) and 80% of the liquid entering the reboiler (L′) is vaporized to give the vapor (V′) leaving the reboiler and returned to the column.

Therefore, at the bottom of the column, if L′ = 1.0 and V′ = 0.8.

The mass balance at the bottom of the column with the reboiler is:

$$L' = V' + B \tag{19.241}$$

or

$$V' = L' - B \tag{19.242}$$

The component balance at the bottom, Lower Operating Line (LOL) is:

$$V' y_m = L' x_{m+1} - B x_B \tag{19.243}$$

or

$$y_m = \frac{L'}{V'} x_{m+1} - \frac{B}{V'} x_B \tag{19.244}$$

$$\text{Slope is } \frac{L'}{V'} = \frac{1}{0.8} = 1.25$$

Using the McCabe–Thiele diagram of Figure 19.51, draw the lower operating line to intercept the “q” line; then draw the upper operating line and extend to the intercept I.

$$I = \frac{x_D}{(R+1)} = 0.22 \text{ (from the McCabe–Thiele diagram)} \tag{19.245}$$

$$\frac{0.975}{(R+1)} = 0.22, \text{ hence}$$

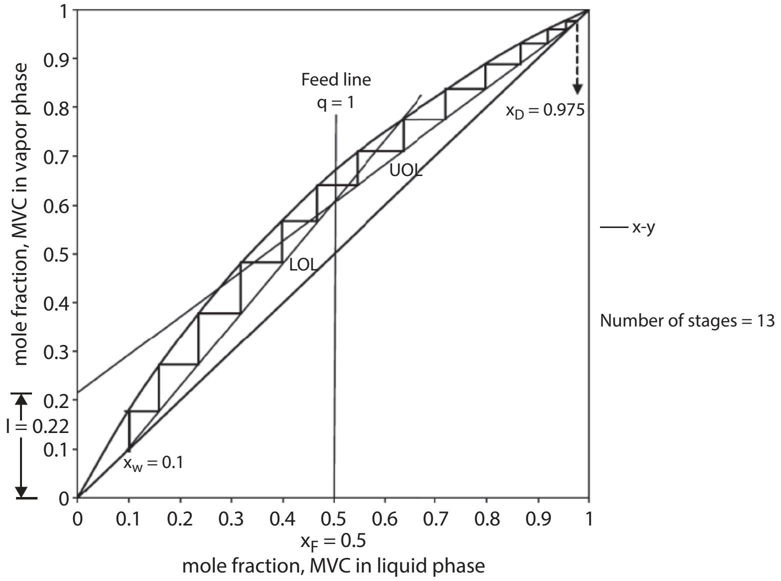


Figure 19.51 The McCabe–Thiele diagram of Example 19.20.

$$R = 3.43$$

From McCabe–Thiele diagram:

$$N + 1 = 14 \text{ stages, hence}$$

$$N = 13 \text{ stages.}$$

R_{min}: from the points (x_D, y_D), a straight line is drawn that touches the VLE line with the "q" line and extended to the y-axis at a point $I_{min} = \frac{x_D}{(R_m + 1)}$

$$\begin{aligned} I_{min} &= \frac{x_D}{(R_m+1)} = 0.34 \\ \frac{0.975}{(R_m+1)} &= 0.34 \\ R_m &= 1.86 \end{aligned} \tag{19.346}$$

(ii) Applying the Underwood Equation to find the minimum reflux ratio:

$$\begin{aligned} R_m &= \frac{1}{(\alpha-1)}\left[\frac{x_{LK,D}}{x_{LK,F}} - \frac{\alpha(1-x_{LK,D})}{(1-x_{LK,F})}\right] \\ &= \frac{1}{(2-1)}\left[\frac{0.975}{0.5} - \frac{2(1-0.975)}{(1-0.5)}\right] \\ &= 1.85 \end{aligned} \tag{19.211}$$

The Fenske method for the minimum number of theoretical stages at total reflux is:

$$\begin{aligned} N_m &= \frac{\ln\left[(x_{LK}/x_{HK})_D (x_{HK}/x_{LK})_B\right]}{\ln(\alpha_{LK}/\alpha_{HK})_{avg}} \\ &= \frac{\ln\left[\left(\frac{0.975}{0.025}\right)\left(\frac{0.9}{0.1}\right)\right]}{\ln 2} = 8.45 \end{aligned} \tag{19.183}$$

From the Gilliland correlation (Figure 19.49), the abscissa axis is:

$$\frac{R - R_m}{R+1} = \frac{3.66 - 1.85}{(3.66+1)} = 0.388 \tag{19.223}$$

The corresponding point on the ordinate axis of the Gilliland correlation is 0.31.

Therefore, the number of theoretical stages can be determined from the ordinate by:

$$\frac{N-N_m}{N+1}=0.31$$

$$\frac{N-8.45}{N+1}=0.31, \text{hence}$$

$$N = 12.69 \text{ (13 stages)}$$

Example 19.21: Graphical Design of a Fractionation Column for Binary Systems

The benzene–toluene example of Robinson and Gilliland [59] has been elaborated on and expanded after the advanced distillation course of Holland [25], Figure 19.52 [59].

It is desired to separate an equimolal mixture of benzene and toluene into a top product containing 95 mol % benzene and a bottom product containing 95 mol % toluene. The distillation is to be carried out at atmospheric pressure. Use a total condenser.

A. Calculate the minimum reflux ratio if the feed is liquid at its boiling point.
B. Calculate the theoretical plates required if a reflux ratio (L/D) of 1.5 times the minimum is employed.

Feed = 50 mols benzene + 50 mols toluene

Overhead = 95% benzene

Bottoms = 95% toluene

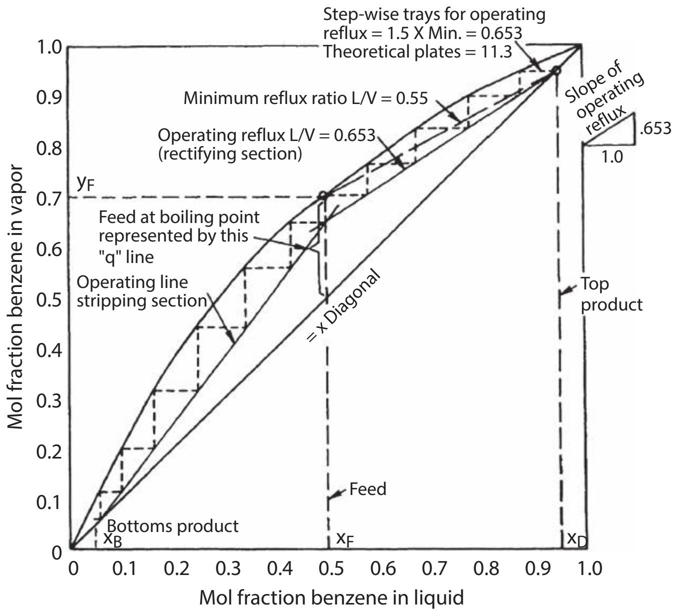


Figure 19.52 Equilibrium curve; benzene–toluene for Example 19.21 (curve data only). Used by permission, Robinson, C. S. R and Gilliland, E. R. Elements of Fractional Distillation, 4th ed., McGraw-Hill Book Co. (1950), All rights reserved.

Solution

Basis: Feed 100 mols/h

Material balance with respect to benzene (more volatile component, MVC)

$$F = D + B$$

Component balance on the MVC

$$Fx_F = D\,x_D + B\,x_B$$

$$0.50\ (100) = (0.95)\ (D) + 0.05B$$

$$50 = 0.95D + 0.05\ (100 - D)$$

$$45 = 0.90D$$

$$D = 45/\ 0.90 = 50 \text{ mols/h overhead product}$$

$$D = \text{overhead product, mols/h}$$

$$B = \text{bottoms, mols/h}$$

A. For a Feed at its Boiling Point:

$$\frac{L_n}{V_n} = \frac{L/D}{L/D+1} \text{ or } \frac{L}{V} = \frac{x_D - y_F}{x_D - x_F},\ y_F = 0.70\,(\text{from curve})$$
$$= \frac{0.95-0.70}{0.95-0.50}$$

Minimum L/V = 0.55 mol reflux/mol vapor up

Substituting: $0.55 = \dfrac{L/D}{L/D+1}$

$$0.55\ L/D + 0.55 = L/D$$

$$0.45\ L/D = 0.55$$

Reflux/Product = L/D = 0.55/0.45 = 1.22

The value of L/D minimum should be equal to:

$$\frac{L_{12}}{D}=\frac{x_D-y_c}{y_c-x_c}=\frac{0.95-0.70}{0.70-0.50}=1.25$$

The slight difference is probably due to inaccuracy in reading $y_c = 0.70$ from equilibrium curve.

B. Number of Theoretical Plates at L/D = 1.5 times minimum:

Operating Reflux Ratio = (1.5) (1.25) = 1.878 = L/D

Slope of operating line at this reflux ratio:

$$\frac{L}{V}=\frac{L/D}{L/D+1}$$

$$\frac{L}{V}=\frac{1.878}{1.878+1}=0.653$$

From Graph, L/V = 0.653 was plotted based on feed at its boiling point, number of theoretical plates (step-wise graph) = 11.3

Now, to calculate theoretical plates:

Rectifying Section

$$y_n=\frac{L_{n+1}}{V_n}x_{n+1}+\frac{D}{V_n}x_D \text{ operating line}$$

At: L/D = 1.878, D = 50 mols/h overhead

L = (1.878) (50) = 93.9 mols/h reflux to column

V = L + D

= 93.9 + 50 = 143.9 mols/h to vapor overhead

$$y_n=\frac{93.9}{143.9}x_{n+1}+\frac{50}{143.9}(0.95)=0.652x_{n+1}+0.331$$

For a total condenser: $y_{top} = x_D = x_R = 0.95$

From the equilibrium curve at $y_t = 0.95$

then:

$$x_t = 0.88$$

$$y_{(t-1)} = 0.651\ (x_t) + 0.331$$

$$y_{(t-1)} = 0.651(0.88) + 0.331 = 0.903$$

$$y_{t-1} = 0.903, \text{ then } x_{t-1} \text{ from equilibrium curve} = 0.788$$

Now calculate y_{t-2}

$$y_{t-2} = 0.651\ (0.788) + 0.331 = 0.844$$

$$\text{At } y_{t-2} = 0.844, \text{ curve reads: } x_{t-2} = 0.69$$

$$\text{Then: } y_{t-2} = 0.651\ (0.69) + 0.331 = 0.780$$

$$\text{At } y_{t-3}, \text{ curve reads: } x_{t-3} = 0.60$$

$$\text{Then: } y_{t-4} = 0.651\ (0.60) + 0.331 = 0.722$$

$$\text{At } y_{t-4}, \text{ curve reads: } x_{t-4} = 0.52 \text{ (Feed Tray)}$$

$$\text{Then: } y_{t-5} = 0.651\ (0.52) + 0.331 = 0.669 \text{ (too far below feed).}$$

Now go to stripping section curve:

Stripping Section

$$y_m = \frac{L_{m+1}x_{m+1}}{V_m} - \frac{W}{V_m}x_B$$

The feed was at its boiling point:

$$V_n = V_m = 143.9$$

$$B = F - D$$

$$= 100 - 50$$

B = Bottoms = 50 mols/h

$$L_m = B + V = 50 + 143.9 = 193.9 \text{ mols/h}$$

$$y_m = \left(\frac{193.9}{143.9}\right) x_{m+1} - \frac{50}{143.9}(0.50)$$

Starting at t – 4 = feed tray:

$$x_{t-4} = 0.52$$

$$y_{(feed-1)} = 1.35\, x_f - 0.0176\,(f - 1)$$

$$y_{(f-1)} = 1.35\,(0.52) - 0.01736 + 0.685$$

$$\text{At } y_{t-1} = 0.685,\ x_{f-1} = 0.475$$

Note: This is not too accurate due to switched operating line equations before the feed compositions were reached, yet, one more calculation on the stripping line would have placed us below the feed plate composition. Hence a change in reflux ratio is necessary in order to split right at the feed composition [81].

Continuing:

$$y_{f-2} = 1.35\,(0.475) - 0.01736 = 0.624$$

From curve at $y_{f-2} = 0.624$

$$x_{f\text{-}2} = 0.405$$

$$y_{f\text{-}3} = 1.35\,(0.405) - 0.01736 = 0.531$$

From curve, $x_{f-3} = 0.32$

$$y_{f-4} = 1.35\,(0.32) - 0.01736 = 0.416$$

$$x_{f-4} = 0.23$$

$$y_{f-5} = 1.35\,(0.23) - 0.01736 = 0.294$$

$$x_{f-5} = 0.15$$

$$y_{f-6} = 1.35\,(0.15) - 0.01736 = 0.186$$

$$x_{f-6} = 0.092$$

$$y_{f-7} = 1.35\ (0.092) - 0.01736 = 0.107$$

$$x_{f-7} = 0.05\ \text{(The desired bottoms composition)}$$

Total number of theoretical trays:
rectifying section = 4
feed tray = 1
stripping section = 7
total = 12 Trays

Example 19.22: Variation of Theoretical Plates in a Column With Thermal Condition of Feed

Using the same operating reflux ratio as in Example 19.21, calculate the theoretical plates required for feed of the following thermal conditions: Use Figure 19.52a.

(a) q = 1.5
(b) q = 0
(c) q = −1.5

A. For q = 1.5

Slope of "q" line = −q/1 − q

Substituting, slope = $\dfrac{-1.5}{1-1.5} = \dfrac{-1.5}{-0.5} = +3$

Referring to calculations of Example 19.21, for an equimolal mixture of benzene and toluene in feed:

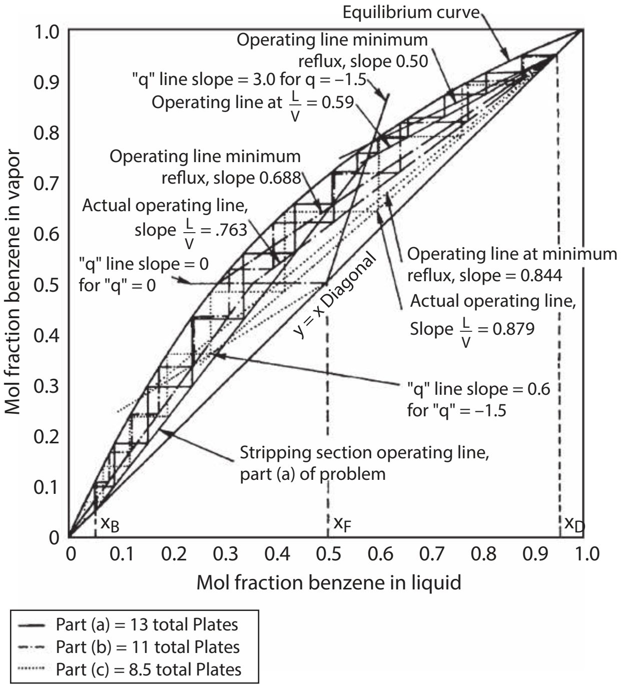


Figure 19.52a Equilibrium curve; benzene–toluene for Example 19.22 (curve data). Used by permission, Robinson, C. S. R and Gilliland, E. R., Elements of Fractional Distillation, 4th ed., McGraw – Hill Book Co., (1950), All rights reserved.

overhead product, D = 50 mols/l00 mols feed

$$\text{calculate}\left(\frac{L_R}{D}\right)_{\text{min}} = \frac{x_D - y_c}{y_c - x_c},$$

where $x_D = 0.95$

$$y_c = 0.774^*$$

$$x_c = 0.59^*$$

$$= \frac{0.95 - 0.774}{0.774 - 0.59} = \frac{0.176}{0.184}$$

$$(L/D)_{\text{min}} = (L_R/D)_{\text{min}} = 0.956 \text{ min. reflux ratio, reflux/product}$$

Slope of Operating Line at Minimum Reflux Ratio:

$$\left(\frac{L}{V}\right) = \left(\frac{L_R}{V}\right)_{\text{min}} = \frac{L/D}{L/D+1} = \frac{0.956}{0.956+1} = 0.49$$

(Graph reads 0.59 but this depends upon accuracy of plot.)

Calculating

$$\left(\frac{L}{V}\right)_{\text{min}} = \frac{x_D - y_c}{x_D - x_c} = \frac{0.95 - 0.774}{0.95 - 0.59} = 0.49$$

Actual Operating Line

Operating reflux ratio = (1.5) (L/D) = 1.5 (0.956)
= 1.432 reflux/product

Slope of actual operating line:

$$\left(\frac{L}{V}\right) = \frac{L/D}{L/D+1} = \frac{1.432}{1.432+1} = 0.59$$

Graphically, read 13 steps or theoretical plates from the top plate through bottom reboiler (assuming a total condenser).

---

*Read from graph at intersection of "q" line for 1.5 and minimum reflux operating line.

rectifying section = 5
feed plate = 1
stripping section = 7 (includes reboiler)
13 Plates including reboiler

To calculate this stepwise:

Operating line of rectifying section:

$$y_{n+1} = \frac{L_r}{V_r} x_n + \frac{D}{V_r} x_D$$

$$L/V = 0.59$$

$$L/D = 1.432, D = 50 \text{ mols product}$$

$$L = (50)\ 1.432 = 71.6 \text{ mols liquid reflux}$$

$$V_r = L_r + D = 71.6 + 50 = 121.6 \text{ mols}$$

then:

$$y_{n+1} = 0.59\ x_n + 50\ (0.95)/121.6$$

$$y_{n+1} = 0.59\ x_n + 0.39, \text{ operating line equation}$$

At top:

$$y_{n+1} = x_D = 0.95$$

So: From equilibrium curve at $y_{n+1} = 0.95$, read the liquid in equilibrium, which is x_n (or top plate in this case) $x_n = x_{top} = 0.88$.

Now substitute this value x = 0.88 into the equation and calculate the vapor coming up from the first plate below the top (t − 1). Thus, if x_n = top plate, y_{n+1} = vapor from plate below top. Now, read equilibrium curve at $y_{(t-1)}$ and get $x_{(n+1)}$ or x_{t-1} which is liquid on plate below top. Then using x_{t-1}, calculate y_{t-2} (second plate below top, etc.). Then, read equilibrium curve to get corresponding liquid x_{t-2}. Continue until feed plate composition is reached, then switch to equation of stripping section and continue similarly until desired bottoms composition is reached.

Operating line of stripping section:

$$y_m = \frac{L_s x_{m+1}}{V_s} - \frac{B}{V_s} x_B$$

Because the feed is a super cooled liquid, $L_s/V_s \neq L_r/V_r$. From definition of "q":

$$L_s = L_r + qF$$

$$L_s = 71.6 + (1.5)\ (100)$$

$$L_s = 221.6$$

Also:
$$\frac{V_r - V_s}{F} = 1 - q \text{ so } \frac{121.6 - V_s}{100} = 1 - 1.5$$

$$121.6 - V_s = -50$$

$$V_s = 171.6$$

Therefore,
$$\frac{L_s}{V_s} = \frac{221.6}{171.6} = 1.29$$

$$\frac{B}{V_s} = \frac{50}{171.6} = 0.291$$

Stripping section operating line:

$$y_m = 1.29\ x_{m+1} - 0.291\ x_B$$

$$x_B = 0.05$$

$$y_m = 1.29\ x_{m+1} - 0.01455$$

Use this equation as described above proceeding down from the feed plate cross-over from the rectifying equation to the stripping equation.

B. For q = 0

This represents feed as all vapor (not superheated). Slope of “q” line:

$$= \frac{-q}{1-q} = \frac{-0}{1-0} = 0$$

This represents no change in overflow from the feed plate, and the increase in vapor flow is equal to the mols of feed.

$$\text{Minimum reflux:} \left(\frac{L}{D}\right)_{\min} = \frac{x_D - y_c}{y_c - x_c},$$

where:

$$X_D = 0.95$$

$$\left.\begin{aligned} y_c &= 0.50 \\ x_c &= 0.29 \end{aligned}\right\} \text{ read from graph}$$

$$= \frac{0.95 - 0.50}{0.50 - 0.29} = 2.14 \text{ min. reflux ratio, reflux/product}$$

Slope of operating line at minimum reflux ratio:

$$\left(\frac{L}{D}\right)_{min} = \frac{L/D}{L/D+1} = \frac{2.14}{2.14+1} = 0.682$$

Slope from graph = 0.688
Operating reflux ratio = (1.5) (2.14) = 3.21, reflux/product, $(L/D)_{op}$
Slope of operating line = $(L/V)_{op}$ = 3.21/(3.21 + 1) = 0.763

No. of theoretical plates from graph	= 11
No. plates rectifying section	= 5
Feed plate	= 1
Stripping section	= 5 (includes reboiler)
Total	= 11 (includes reboiler)

Rectifying Section Equation for Operating Line

$$y_{n+1} = \frac{L_r}{V_r} x_n + \frac{D}{V_r} x_D$$

$$L/V = 0.763$$

$$L/D = 3.21$$

L = (3.21) (50 mol product. D) = 160.6 mols (reflux liquid)

$$V_r = L_r + D = 160.6 + 50$$

V_r = 210.6 mols vapor up column

then:

$$y_{n+1} = 0.763 x_n + \frac{50}{210.6}(0.95)$$

$$y_{n+1} = 0.763\ x_n + 0.225$$

Liquid Down Stripping Section

$$L_s = L_r + qF$$

$$L_s = 160.6 + (0)\ (100 \text{ mols feed})$$

$$L_s = 160.6 = L_r$$

Vapor Up Stripping Section

$$\frac{V_r - V_s}{F} = 1 - q$$

$$\frac{210.6 - V_s}{100} = 1 - 0$$

$$210.6 - V_s = 100$$

$$V_s = 110.6 \text{ mols}$$

Stripping Section Equation for Operating Line

$$y_m = \frac{L_s}{V_s} x_{m+1} - \frac{B}{V_s} x_B$$

$$y_m = \frac{160.6}{110.6} x_{m+1} - \frac{50}{110.6}(0.05)$$

$$y_m = 1.452\ x_{m+1} - 0.0226$$

Use these equations as described for the (a) part of problem in solving for number of theoretical plates stepwise.

C. For $q = -1.5$

This represents feed as a superheated vapor, and there is a decrease in liquid overflow from feed plate.

Slope of "q" line = $\frac{-q}{1-q} = \frac{-(-1.5)}{1-(-1.5)} = 0.60$

Minimum reflux ratio: $(L/D)_{min} = \frac{x_D - y_c}{y_c - x_c}$

where

$$\left.\begin{array}{l} x_D = 0.95 \\ y_c = 0.277 \\ x_c = 0.138 \end{array}\right\} \text{read from graph}$$

$$= \frac{0.95 - 0.277}{0.277 - 0.138}$$

$$(L/D)_{min} = 4.84 \text{ reflux/product}$$

Slope of operating line at minimum reflux ratio:

$$\left(\frac{L}{V}\right)_{min} = \frac{L/D}{L/D + 1} = \frac{4.84}{4.84 + 1} = 0.830(\text{graph reads } 0.844)$$

Actual Operating Line

Operating reflux ratio = (1.5) (4.84) = 7.26 reflux/product

Slope of actual operating line = (L/D) = $\frac{7.26}{7.25 + 1} = 0.879$

Graphically we read 8.5 total plates through to the bottom reboiler

rectifying section	= 5
feed plate	= 1
stripping section	= 2.5 (includes reboiler)
total	= 8.5 (includes reboiler)

Equations for Stepwise Tray to Tray Calculations Rectifying Section Operating Line

$$y_{n+1} = \frac{L_r}{V_r} x_n + \frac{D}{V_r} x_D$$

$$L_r/V_r = 0.879$$

$$L/D = 7.26$$

$$L_r = (7.26)\ (50) = 363 \text{ mols liquid reflux}$$

$$V_r = L_r + D = 363 + 50 = 413$$

$$y_{n+1} = 0.879 x_n + \frac{50}{413}(0.95)$$

$$y_{n+1} = 0.879x_n + 0.115$$

Liquid Down Stripping Section

$$L_s = L_r + qF$$

$$L_s = 363 + (-1.5)(100) = 213 \text{ mols liquid}$$

Vapor Up Stripping Section

$$\frac{V_r - V_s}{F} = 1 - q$$

$$\frac{413 - V_s}{100} = 1 - (-1.5) = 2.5$$

$$-V_s = 250 - 413$$

$$V_s = 163 \text{ mo1s vapor}$$

Stripping Section Operating Line

$$y_m = \frac{L_s}{V_s} x_{m+1} - \frac{B}{V_s} x_B$$

$$y_m = \frac{213}{163} x_{m+1} - \frac{50}{163}(0.05)$$

$$y_m = 1.307x_{m+1} - 0.01535$$

Use these equations as described for Part (a) in solving for theoretical plates.

19.31 Determination of Number of Theoretical Plates in Fractionating Columns by the Smoker Equations at Constant Relative Volatility (α = constant)

Smoker [100] developed an analytical equation to determine the number of stages when the relative volatility, α is constant. This method can be used in preference to the McCabe–Thiele [81] graphical method because the accuracy of the latter depends on the care given to the construction of the plots. The graphical construction for the number of stages also becomes more difficult at very low concentrations. Further if the relative ≃1.0, the number of stages required will be very large, and the diagram will be difficult to construct.

The Equations

Smoker's equations consider Raoult's law for a binary mixture, where α is a constant and represent the vapor–liquid equilibrium relationship. From Equation 19.48

$$y = \frac{\alpha x}{1 + (\alpha - 1)x} \tag{19.48}$$

Any operating line that is either above or below the feed plate can be represented by the general equation of a straight line:

$$y = mx + b \tag{19.247}$$

Eliminating y from Equations 19.48 and 19.247 gives a quadratic equation in x

$$m(\alpha - 1)x^2 + [m + (\alpha - 1)b - \alpha]x + b = 0 \tag{19.248}$$

For any particular distillation problem, Equation 19.248 will have only one real root k, between 0 and 1. Representing this root k, and substituting in Equation 19.248 gives

$$m(\alpha - 1)k^2 + [m + (\alpha - 1)b - \alpha]k + b = 0 \tag{19.249}$$

k is the value of the x-ordinate at the point where the extended operating lines intersect the vapor–liquid equilibrium curve.

Smoker shows that the number of stages required is given by

$$N = \log\left[\frac{x_o^*\left(1 - \beta x_n^*\right)}{x_n^*\left(1 - \beta x_o^*\right)}\right] \Big/ \log\left(\frac{\alpha}{mc^2}\right) \tag{19.250}$$

where

$$\beta = \frac{mc(\alpha - 1)}{\alpha - mc^2} \tag{19.251}$$

N = number of stages required effecting the separation represented by the concentration change from

$$x_n^* \text{ to } x_o^*; x^* = (x - k) \text{ and } x_o^* > x_n^*; c = 1 + (\alpha - 1)k \tag{19.252}$$

k = composition of the liquid where the operating line intersects the equilibrium line
m = slope of the operating line between x_n^* and x_o^*
α = average relative volatility, assumed constant over x_n^* to x_o^* = $(\alpha_{top} + \alpha_{bottom})/2$

19.31.1 Application of Smoker's Method to a Binary Distillation Column

Applying Smoker's equation to actual distillation conditions with a single feed and no side streams.

Rectifying Section

For a rectifying column, the number of plates required to enrich from the feed composition, x_F to the distillate (product) composition, x_D at the fixed reflux ratio, R is obtained from:

$$m = \frac{R}{R+1} \tag{19.253}$$

and

$$b = \frac{x_D}{R+1} \tag{19.254}$$

Also

$$m(\alpha - 1)k^2 + [m + (\alpha - 1)b - \alpha]k + b = 0 \tag{19.255}$$

where k is the root between 0 and 1

$$x_o = x_D \tag{19.256}$$

$$x_o^* = x_D - k \tag{19.257}$$

$$x_n = x_F \tag{19.258}$$

$$x_n^* = x_F - k \tag{19.259}$$

Stripping Section

For the stripping section, the bottoms composition, x_B, the feed, x_F, the overhead composition, x_D, and the reflux ratio, R all fix the operating line. The values of m and b are:

$$m = \frac{Rx_F + x_D - (R+1)x_B}{(R+1)(x_F - x_B)} \tag{19.260}$$

and

$$b = \frac{(x_F - x_D)x_B}{(R+1)(x_F - x_B)} \tag{19.261}$$

$$x_o^* = x_F - k \tag{19.262}$$

$$x_n^* = x_B - k \tag{19.263}$$

If the feed stream is not introduced at its bubble point, x_F is replaced by the value of x at the intersection of operating lines given by:

$$x_F^* = \frac{b + x_F/(q-1)}{q/(q-1) - m} \tag{19.264}$$

For distillation at total reflux R = ∞, m = 1, b = 0, k = 0, c = 1 and Equation 19.250 becomes:

$$N = \log\left\{\frac{x_o(1-x_n)}{x_n(1-x_o)}\right\} \Big/ \log\alpha \tag{19.265}$$

Equation 19.250 simplifies to the equation derived by Underwood [88]. Smoker's equation is not limited to the range of low concentrations, but can be used for stage calculation over the entire range. In such instances, the equation is written twice, one for each section of the tower. The average relative volatilities for the rectifying section and for the stripping section are determined.

Example 19.23: Use of Smoker's Equation to Determine the Number of Theoretical Stages and The Optimum Feed Stage for Binary Distillation

A column is to be designed to separate 1000 moles/h of a binary mixture of benzene (C_6H_6) and toluene (C_7H_8). The feed will contain 40% benzene and 60% toluene. A distillate that is 99% benzene and a bottoms that is 1% benzene are desired at a reflux ratio of 3 to 1. For this mixture, the average value of relative volatility (α) is 2.50. Estimate the number of equilibrium stages at this reflux ratio and the optimum feed stage location.

Solution

$$x_F = 0.4,\ x_D = 0.99,\ x_B = 0.01,\ R = 3.0,\ \alpha = 2.5$$

Rectifying Section

$$m = \frac{R}{R+1} = \frac{3}{3+1} = 0.75$$

$$b = \frac{x_D}{R+1} = \frac{0.99}{3+1} = 0.2475$$

$$m(\alpha - 1)k^2 + [m + (\alpha - 1)b - \alpha]k + b = 0 \tag{19.255}$$

$$0.75(2.5 - 1)k^2 + [0.75 + 0.2475(2.5 - 1) - 2.5]k + 0.2475 = 0$$

Solving: $1.125k^2 - 1.37875k + 0.2475 = 0$

$$k = \frac{-(-1.37875) \pm \sqrt{(-1.37875)^2 - 4(1.125)(0.2475)}}{2 \times 1.125}$$

$k = 1.0071$ or 0.21847

Since $0 < k < 1$, Use $k = 0.21847$

$$x_o^* = x_D - k = 0.99 - 0.2185 = 0.7715$$

$$x_n^* = x_F - k = 0.4 - 0.2185 = 0.1815$$

$$c = 1 + (\alpha - 1)k = 1 + (2.5 - 1)0.2185 = 1.328$$

$$\beta = \frac{mc(\alpha - 1)}{\alpha - mc^2} = \frac{0.75 \times 1.328(2.5 - 1)}{2.5 - 0.75 \times 1.328^2} = 1.268$$

$$N = \log\left[\frac{x_o^*\left(1 - \beta x_n^*\right)}{x_n^*\left(1 - \beta x_o^*\right)}\right] \Big/ \log\left(\frac{\alpha}{mc^2}\right)$$

$$\begin{aligned} N &= \log\left[\frac{0.7715(1 - 1.268 \times 0.1815)}{0.1815(1 - 1.268 \times 0.7715)}\right] \Big/ \log\left(\frac{2.5}{0.75 \times 1.328^2}\right) \\ &= 7.88 \text{ stages} \end{aligned} \tag{19.250}$$

Stripping Section

Feed is taken at its bubble point.

$$\begin{aligned}
m &= \frac{Rx_F + x_D - (R+1)x_B}{(R+1)(x_F - x_B)} \\
&= \frac{3\times 0.4 + 0.99 - (3+1)0.01}{(3+1)(0.4-0.01)} = 1.3782 \\
b &= \frac{(x_F - x_D)x_B}{(R+1)(x_F - x_B)} = \frac{(0.4-0.99)0.01}{(3+1)(0.4-0.01)} = -0.00378
\end{aligned} \tag{19.260}$$

$$m(\alpha - 1)k^2 + [m + (\alpha - 1)b - \alpha]k + b = 0 \tag{19.255}$$

$$1.3782(2.5 - 1)k^2 + [1.3782 + (2.5 - 1)(-0.00378) - 2.5]k - 0.00378 = 0$$

Solving $$2.0673k^2 - 1.12747k - 0.00378 = 0$$

$$\begin{aligned}
k &= \frac{-(-1.12747) \pm \sqrt{(-1.12747)^2 - 4(2.0673)(-0.00378)}}{2\times 2.0673} \\
&= 0.5487 \text{ or } -0.00333
\end{aligned}$$

Since $0 < k < 1$, use $k = 0.5487$

$$x_o^* = x_F - k = 0.4 - 0.5487 = -0.1487$$

$$x_n^* = x_B - k = 0.01 - 0.5487 = -0.5387$$

$$c = 1 + (\alpha - 1)k = 1 + (2.5 - 1)0.5487 = 1.823$$

$$\begin{aligned}
\beta &= \frac{mc(\alpha-1)}{\alpha - mc^2} = \frac{1.3782\times 1.823(2.5-1)}{2.5 - 1.3782\times 1.823^2} = -1.81168 \\
N &= \log\left[\frac{x_o^*\left(1-\beta x_n^*\right)}{x_n^*\left(1-\beta x_o^*\right)}\right] \Big/ \log\left(\frac{\alpha}{mc^2}\right) \\
&= \log\left[\frac{-0.1487\{1-(-1.81168)(-0.5387)\}}{-0.5387\{1-(-1.81168)(-0.1487)\}}\right] \Big/ \log\left(\frac{2.5}{1.3782\times 1.823^2}\right) \\
&= 7.79
\end{aligned}$$

Table 19.26 Results for calculating equilibrium number of stages using smoker's equation.

Equilibrium Number of Stages by Smoker's Equation	
The Relative Volatility of Component 1.	2.5
The Relative Volatility of Component 2.	2.5
The Mole Fraction of the Feed	0.4
The Mole Fraction of the Distillate	0.99
The Mole Fraction of the Bottoms	0.01
The Reflux Ratio	3
The Number of Stages in the Rectifying Section	7.88
The Number of Stages in the Stripping Section	7.79
Total Number of Equilibrium Stages	16
Optimum Feed Stage location	8

Total number of stages is therefore:

$$N_s = 7.88 + 7.79 = 15.67 \text{ say } 16$$

An Excel spreadsheet program Example 19.23.xlsx has been developed, which uses a quadratic equation to determine the number of stages in the rectifying and stripping sections of a distillate column. The program uses Smoker's equations involving two components to calculate the required number of stages. Table 19.26 shows the results of the calculations of Example 19.23. The required number of stages between benzene and toluene mixture is 16 and the optimum feed location is 8. Constructing the McCabe–Thiele diagram using Figure 19.42a for the more volatile component (i.e., benzene) with the feed at its bubble point (q = 1), gives the number of stages as 18.

19.32 The Jafarey, Douglas, and McAvoy Equation: Design and Control

Operability and control are important considerations for any column [101, 102]. For example, it is common practice for operators to increase the reflux rate above its design value and thereby increase the purity of the overhead stream (assuming that the desired product arises from the overhead). Using this approach, the disturbances in the feed composition seldom cause the top product to fall below its quality specification. However, this advantage must be balanced by the excess steam supply needed to over reflux. To solve problems of this type, and to develop better control systems for columns, Shrinskey [103] developed an empirical correlation between the boilup ratio and the separation factor for a column, and Smoker [100] developed an exact analytical solution for binary distillation columns for the case of constant relative volatility.

Jafarey *et al.* [101, 102] derived an approximate equation for binary distillation by simplifying the solution of Smoker's equation. This equation can be used to predict the effect of disturbances on column performance, and is thus useful in computer and microprocessor control, where it can be applied to estimate the effect of disturbances and the control action required to compensate for them. The equation is:

$$N = \frac{\ln S}{\ln\left[\alpha\sqrt{1-\frac{R+q}{(R+1)(Rz+q)}}\right]} \tag{19.256}$$

where

N = Number of stages.

$$S = (x_{LK}/x_{HK})_D(x_{HK}/x_{LK})_B \tag{19.257}$$

R = Reflux ratio.

q = Number of pound-moles of liquid formed on the feed stage when introducing 1 lb-mole of feed.
z = Mole fraction in the feed.
α = Relative volatility.

Example 19.24: Design and Control of a Distillation Column Using the Jafarey, Douglas, and McAvoy Algorithm

A benzene–toluene column normally operates as described in Example 19.23 with 30% vaporized. The algorithm manipulates boilup flow to control toluene purity. If the toluene purity is to be temporarily, increased from 90 to 95% and the benzene purity is to remain unaffected, to what the boilup flow rate to set? Assume that any boilup changes will be compensated for by reflux and distillate rate changes, such that the benzene purity remains unaffected. The relative volatility is 2.5.

Solution

The overall mass balance:

$$F = D + B$$

$$1000 = D + B$$

Component balance on the MVC (benzene)

$$F_{xF} = D_{xD} + Bx_B$$

$$1000\ (0.4) = D \times 0.99 + B \times 0.1$$

$$400 = 0.99\ D + 0.1\ B$$

$$400 = 0.99\ (1000 - B) + 0.1B$$

$$400 = 990 - 0.99B + 0.1B$$

$$B = 663 \text{ lb mole/h}$$

$$D = 1000 - 663$$

$$= 337 \text{ lb mole/h}$$

In the rectifying section of the column

$$V = L + D$$

But the reflux ratio, R is:

$$R = L/D$$

$$L = R \times D = 3 \times 337$$

$$= 1011 \text{ lb mole/h}$$

Mass balance in the rectifying section:

$$V = L + D$$

$$= 1011 + 337$$

$$= 1348 \text{ lb-mole/h}$$

Component balance in the rectifying section of the column

$$V y_n = L_{n+1} x_{n+1} + D x_D$$

$$y_n = \frac{L_{n+1}}{V_n} x_{n+1} + \frac{D}{V_n} x_D$$

$$= \frac{1011}{1348} x_{n+1} + \frac{337}{1348} x_D$$

$y_n = 0.75x_{n+1} + 0.25\ x_D$ passes through (0.99, 0.99)

At the feed stage, since 30 percent of the feed is vaporized, q = 0.7. L′ from the definition is:

$$
\begin{aligned}
L' &= L + qF \\
&= 1011 + 0.7\ (1000) \\
&= 1011 + 700 \\
L' &= 1711\ \text{lb} - \text{mole/h}
\end{aligned}
$$

In the stripping section of the column:

$$V' = L' - B$$

$$= 1711 - 663$$

$$= 1048\ \text{lb-mole/h}$$

The component balance in the stripping section of the column

$$
\begin{aligned}
V'_m y_m &= L'_{m-1} x_{m-1} - Bx_B \\
y_m &= \frac{L'_{m-1}}{V'_m} x_{m-1} - \frac{B}{V'_m} x_B \\
&= \frac{1711}{1048} x_{m-1} - \frac{663}{1048} x_B
\end{aligned}
$$

$y_m = 1.6323\ x_{m-1} - 0.6322x_B$ passes through (0.1, 0.1)

Slope of the q-line is:

$$\frac{q}{(q-1)} = \frac{0.7}{(1-0.7)} = -2.33$$

From Equation 19.256 is:

$$N = \frac{\ln S}{\ln\left[\alpha\sqrt{1 - \frac{R+q}{(R+1)(Rz+q)}}\right]}$$

where

$$S = \left(\frac{x_{LK}}{x_{HK}}\right)_D \left(\frac{x_{HK}}{x_{LK}}\right)_B$$

$$\begin{aligned}\ln S &= \ln\left[\left(\frac{0.99}{0.01}\right)\left(\frac{0.90}{0.10}\right)\right] \\ &= 6.79 \\ N &= \frac{6.79}{\ln\left[2.5\sqrt{1-\frac{3+0.7}{(3+1)(3\times0.4+0.7)}}\right]} \\ &= 11.65 \text{ trays.} \\ &= 12 \text{ trays}\end{aligned}$$

Solve the material and component balances for the new conditions

$$1000 = D + B$$

$$1000 \text{ x } 0.4 = D \text{ x } 0.99 + 0.05B$$

$$400 = 0.99D + 0.05B$$

Solving these simultaneously gives B = 628 lb-mole/h and D = 372 lb-mole/h.

Recalculate S for the new conditions:

$$\begin{aligned}\ln S &= \ln\left[\left(\frac{0.99}{0.01}\right)\left(\frac{0.95}{0.05}\right)\right] \\ &= 7.54\end{aligned}$$

$$N = \frac{\ln S}{\ln\left[\alpha\sqrt{1-\frac{R+q}{(R+1)(Rz+q)}}\right]}$$

$$\frac{\ln S}{N} = \ln\left[\alpha\sqrt{1-\frac{R+q}{(R+1)(Rz+q)}}\right]$$

$$\frac{\ln S}{N} = \frac{7.54}{12} = 0.628$$

$$e^{0.628} = \left[\alpha\sqrt{1 - \frac{R+q}{(R+1)(Rz+q)}}\right]$$

$$\left(\frac{1.8744}{2.5}\right)^2 = 1 - \frac{R+q}{(R+1)(Rz+q)}$$

$$0.562 = 1 - \frac{R+1}{(R+1)(Rz+q)}$$

or

$$\frac{R+q}{(R+1)(Rz+q)} = 1 - 0.562 = 0.438$$

$$\frac{R+0.7}{(R+1)(0.4R+0.7)} = 1 - 0.562 = 0.438$$

$$R + 0.7 = 0.438[(R + 1)(0.4R + 0.7)]$$

This gives the quadratic equation:

$$0.175R^2 - 0.5184R - 0.3935 = 0$$

Solving for R gives

$$R = \frac{0.5184 \pm \sqrt{0.5184^2 + 4 \times 0.175 \times 0.39}35}{2 \times 0.175}$$
$$= 3.58$$

Since D = 372 lb-mole/h, L = 3.58 × 372 = 1332 lb − mole/h

$$L' = L + qF$$

$$= 1332 + 0.7 \times 1000 = 2032 \text{ lb-mole/h.}$$

$$V' = L' - B$$

$$= 2032 - 628 = 1404 \text{ lb-mole/h}$$

Therefore, the computer control will therefore require increasing the boil-up rate from 1048 lb-mole/h to 1404 lb-mole/h.

Summary

Variable	**This variable gives a measure of**	**lb-mole/h**
L	Rectifying section liquid load	1011
V	Rectifying section vapor load condenser duty	1348
L′	Stripping section liquid load	2032
V′	Stripping section vapor load reboiler duty.	1404

Example 19.25: Calculation of Minimum Number of Theoretical Trays/Plates/Stages at Total Reflux for a Binary System

A finishing column is required to produce trichlorethylene at 99.9% (vol.) purity from 10,000 lb/h of a feed of 40% (wt.) trichlorethylene and 60% (wt.) perchlorethylene. Only 1% (vol.) of the trichlorethylene can be accepted in the bottoms.

Because the process system that will receive vents from his condensing system is operating at 5 psig, allow 5 psi pressure drop to ensure positive venting and set top of tower pressure at 10 psig.

Feed (158°F)	**(1) Wt %**	**(2) Mol Wt**	**(1) (2) Mols**	**Mol fraction**
Trichlorethylene	40	131.4	0.00304	0.456
Perchlorethylene	60	165.9	0.00362	0.544
Total	100		0.00666	1.000

Avg mol wt 1.00/0.00666 = 150.0

Overhead

Overhead temperature for essentially pure products at 10 psig = 223°F from vapor pressure curve.

Bottoms

Allow 10 psi tower pressure drop, this makes bottom pressure = 20 psig = 1800 mm Hg.

Material Balance:

$$F = D + B$$

$$\text{Feed rate: mols/h} = \frac{10{,}000 \text{ lb/h}}{(150.0)} = 66.7$$

$$66.7 = D + B$$

or

$$B = 66.7 - D$$

Component balance:

$$F\,x_1 = D\,x_1 + B\,x_1$$

$$(66.7)\,(0.456) = D\,(0.999) + B\,(0.01)$$

$$30.4 = 0.999\,D + 0.01\,(66.7 - D)$$

$$D = 30.05 \text{ mols/h}$$

$$\text{bottoms: } B = F - D$$

$$B = 66.7 - 30.05 = 36.65 \text{ mols/h}$$

Bottoms compositions	**Mol fraction**	**Mols/h**	**V.P. (316°F) mm Hg**	**V.P. mm Hg**
Trichlorethylene	0.01	0.3665	4200	42
Perchlorethylene	0.99	36.2835	1780	1762
	1.00	36.65		1804 mm

The 1804 mm compares to the balance value of 1800 mm ≅ 20 psig.

Overhead composition	**Mol fraction**	**Mols/h**
Trichlorethylene	0.999	30.02
Perchlorethylene	0.001	0.03
	1.000	

Relative Volatility: Overhead Conditions

$$\alpha \text{ tri/per}(223°\text{F}) = \frac{\text{v.p.(tri)}}{\text{v.p.(per)}} = \frac{1{,}280\text{mm}}{385} = 3.32$$

Bottom Conditions:

$$\alpha \text{ tri/per}(316°\text{F}) = \frac{\text{v.p.(tri)}}{\text{v.p.(per)}} = \frac{4{,}200\text{mm}}{1{,}780} = 2.36$$

Thermal Condition of the Feed at 158°F

At conditions of feed tray, assume pressure is 15 psig ≅ 1533 mm Hg. Determine bubble point:

Component	x_{iF}	**Assume t = 266°F V.P., mm Hg**	**Partial Press. = V.P. × x_{iF}, mm Hg**
Trichlorethylene	0.456	2350	1072
Perchlorethylene	0.544	880	478
Total			1550

This is sufficiently close to 1533 mm for practical purposes; the actual temperature might be 265°F, although plotted data are probably not that accurate. Because the feed enters at 158°F and its bubble point is 266°F, the feed is sub-cooled.

Heat to vaporize one mol of feed.

Component	x_{iF}	**Latent Ht. @ 266°F Btu/mol**	**(xiF) (Lv) Btu/mol**	**(°F) Cp @ 158°F Btu/mol**	**(x_{iF})(Cp) (266-158) Btu/mol**
Trichlorethylene	0.456	12,280	5600	30.9	1523
Perchlorethylene	0.544	14,600	7950	36.4	2180
			13,550		3703

$$q = \frac{\text{heat required to vaporize one mol of feed}}{\text{latent heat of one mol of feed}}$$

$$q = \frac{13{,}550 + 3{,}703}{13{,}550} = \frac{17{,}253}{13{,}550} = 1.272$$

Minimum Number Tray at Total Reflux

$x_{D1} = 0.999$
$x_{Dh} = 0.001$
$x_{B1} = 0.01$
$x_{BH} = 0.99$
$\alpha_{avg} = 2.8$

For a total condenser system:

$$(N_{min}+1)=\frac{\log\left[(x_{Dl}/x_{Dh})(x_{Bh}/x_{Bl})\right]}{\log\alpha_{avg}}$$

$$=\frac{\log[(0.999/0.001)(0.99/0.01)]}{\log 2.8}=11.17$$

$$N_{min} + 1 = 11.17$$

$$N_{min} = 10.17 \text{ trays, not including reboiler}$$

Summary

Minimum total physical trays in the column	10.17
Reboiler	1.0
For conservative design, add feed tray	1.0
Minimum number of total theoretical stages	12.17, say 12

Minimum Reflux Ratio

Because this is not feed at its boiling point, but sub-cooled liquid, the convenient charts cannot be used with accuracy. Using Underwood's general case:

$$\frac{(L/D)(x_{1F})+qx_{1D}}{(L/D)(1-x_{1F})+q(1-x_{1D})}=\frac{\alpha\{[(L/D)+1]y_{1F}+(q-1)(x_{1D})\}}{[(L/D)+1](1-x_{1F})+(q-1)(1-x_{1D})} \tag{19.258}$$

Solve first for y_{1F}, assuming that the system follows the ideal (as it closely does in this instance).

$$y_{1F}=\frac{x_{1F}(\alpha_F)}{1+(\alpha_F-1)x_{1F}} \tag{19.48}$$

This replaces drawing the equilibrium curve and solving graphically, and is only necessary since the "q" is not 1.0 or 0.

The α should be for the feed tray. However, the value of α = 2.8 should be accepted for feed tray conditions (not 158°F). It would not be if this were predominantly a rectifying or a stripping operation.

$$y_{1F}=\frac{0.456(2.8)}{1+(2.8-1)(0.456)}=0.70$$

Now, substituting to solve for $(L/D)_{min}$

$$\frac{(L/D)(0.456)+1.272(0.999)}{(L/D)(1-0.456)+1.272(1-0.999)} = \frac{2.8\{[(L/D)+1]0.70+(1.272-1)(0.999)\}}{[(L/D+1)(1-0.456)+(1.272-1)(1-0.999)]}$$

$$\frac{(L/D)(0.456)+1.272}{(L/D)(0.544)+0.00127} = \frac{2.8\{(L/D)(0.70)+0.70+0.271\}}{[(L/D)(0.544)+0.544+0.000271]}$$

Solving this quadratic:

$$(L/D)_{min} = 0.644$$

Reading Figure 19.40 for $(L/D)_{min}$ assuming a liquid feed at the boiling point, $(L/D)_{min}$ = 1.2. This demonstrates the value of taking the thermal condition of the feed into account.

Actually, any point on one of the curves represents a condition of reflux and number of trays that will perform the required separation.

19.33 Number of Theoretical Trays at Actual Reflux

Assume actual reflux ratios of 1.2, 1.8, 2.25, 3.0 times the minimum and plot the effect on the number of theoretical plates using Gilliland plot.

Actual Reflux Ratio	$\frac{(L/D)-(L/D)_{min}}{(L/D)+1}$	**(From Fig. 19.49a) $N - N_{min}$**	**Number of Theoretical plates**	**Conservative Add 1 for Feed Total number**
0.772	0.0722	0.552	26.2	27
1.16	0.239	0.416	19.8	21
1.45	0.329	0.356	17.9	19
1.93	0.439	0.288	16.1	17

$$\frac{(L/D)-(L/D)_{min}}{(L/D)+1} = \frac{0.772-0.644}{0.772+1} = 0.0722$$

Read value from curve Figure 19.49a.

$$\frac{N-N_{min}}{N+1} = 0.552, \text{ so } \frac{N-11.18}{N+1} = 0.552$$

$$N = 26.2$$

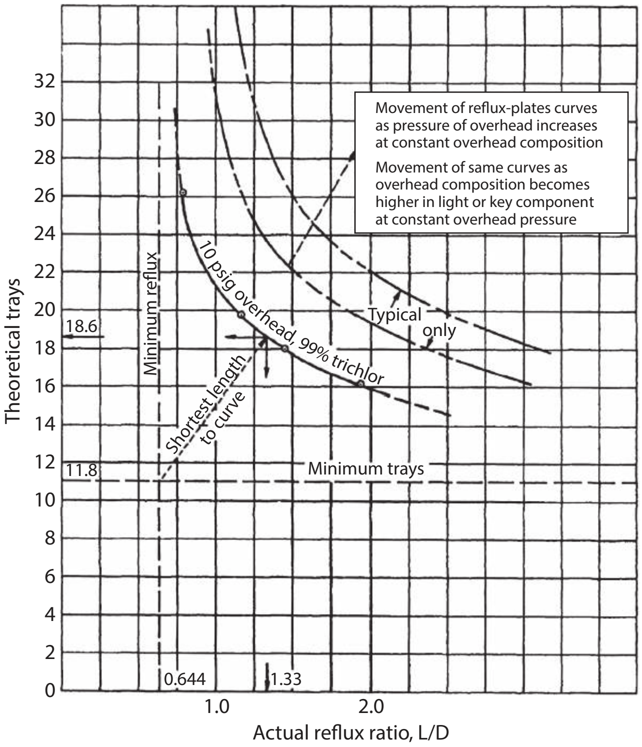


Figure 19.53 Relationship of reflux ratio and theoretical trays, for Example 19.25.

Note that these values are for theoretical trays and as in earlier calculations, require conversion into actual number of trays. Efficiencies are generally in the range of 50–60% for systems of this type. This indicates a column of actual number of trays almost twice the theoretical number at the operating reflux.

Figure 19.53 presents the usual determination of optimum or near optimum theoretical trays at actual reflux based on performance. It is not necessarily the point of least cost for all operating costs, fabrication costs or types of trays. A cost study should be made to determine the merits of moving to one side or other of the so-called optimum point. From the Figure 19.53

First choice, actual reflux ratio, L/D = 1.33

Corresponding theoretical number of trays or stages, N = 18.6

Note that the 18.6 includes the reboiler, so physical trays in column = 17.6. Do not round-off decimal or fractions of trays until after efficiency has been included.

Tray Efficiency

Base at average column temperature of (158 + 266)/2 = 212°F.

Component	x_{iF}	**Viscosity, cP**	**(μ) (x_{iF}), cP**
Trichlorethylene	0.456	0.27	0.123
Perchlorethylene	0.544	0.36	0.196
			0.319

From Figure 19.54:
Efficiency = 47.5%

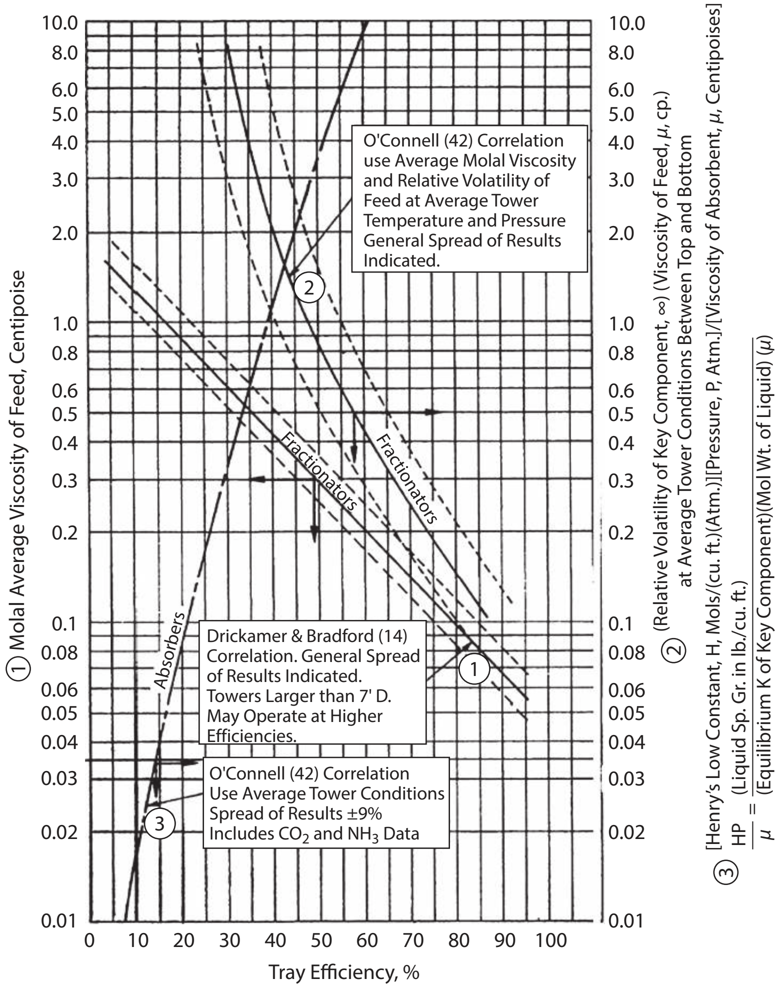


Figure 19.54 Empirical correlations of overall efficiencies for fractionation and absorption.

Actual Trays at Actual Reflux

Actual L/D = 1.33
Actual trays = 18.6/0.475 = 39.2 (including reboiler)
Physical trays: 39.2 – 1 (reboiler) + 1 (conservative, feed) = 39.2
Round-off to: 40 trays plus reboiler plus total condenser.

Note: If there is any reason to know that the efficiency of this system is usually lower (or in same chemical family), then either the efficiency should be reduced to account for this or extra number of trays should be added. In practice, the installed column might contain 40 to 50 trays [81].

Types of Tray

Tray details will be considered in a later example.

19.34 Estimating Tray Efficiency in a Distillation Column

Several empirical efficiency correlations have been developed from commercial equipment and some laboratory data and serve most of design problems for the average hydrocarbon and chemical system. They are empirical correlations, and hence, application in new systems is unpredictable. For this reason, results for efficiencies should be evaluated by several methods to obtain some idea of the possible spread. In light of the American Institute of Chemical Engineers (AIChE) study discussed below, some of these methods can be off the range by 15–50%. Comparisons indicate these deviations are usually on the safe or low side. The relation of Drickamer and Bradford [104] of Figure 19.54 has been found to give good agreement for hydrocarbons, chlorinated hydrocarbons, glycols, glycerine and related compounds, and some rich hydrocarbon absorbers and strippers.

The relation of O'Connell [105] (Figure 19.54) has generally also given good results for the same systems, but generally the values are high. The absorber correlation of O'Connell (Figure 19.54) can be used as long as it generally gives lower values than the other two relations. It can be used for stripping of gases from rich oils provided that great caution is exercised.

The area of absorption and stripping is difficult to correlate for the wide range of peculiarities encountered in such systems. The correlation of Gautreaux and O'Connell [106] allows a qualitative handling of tray mixing to be considered with overall and local efficiencies. In general, it agrees with the Drickamer correlation at least for towers up to 7 ft. (2.1 m) in diameter. Although the effect of liquid path must be apparently considered, the wide variety of tray and cap designs makes this generally possible, and the overall correlations appear to serve adequately.

The American Institute Chemical Engineer's Distillation Tray Efficiency Research [107] program has produced a more detailed method than the shortcut methods and correspondingly is believed to produce reliable results. This method produces information on tray efficiencies of new systems without experimental data. At present, there is not enough experience with using this method and its results to evaluate its complete range of application.

Murphree [108] developed "point" and "overall" distillation tray efficiencies, which are examined in detail in Reference 2. The expressions are [85] as follows:

$$\text{Plate/Tray Efficiency: } E_{MV}^{o} = \frac{y_i - y_o}{y_i - y_e^{*}} \tag{19.259}$$

The plate/tray efficiency is the integrated effect of all the point efficiencies.

$$\text{Point Efficiency: } E_{MV}^{o} = \frac{y_i' - y_o'}{y_i' - y_e} \tag{19.260}$$

$$\text{Overall tray efficiency, } E_o = \frac{\text{Number Theoretical Trays}}{\text{Number of Actual Trays}} \tag{19.261}$$

where

y_i = average composition of vapor entering tray
y_o = average composition of vapor leaving tray
y_e^{*} = composition of vapor in equilibrium with liquid flowing to plate below
y_i' = vapor composition entering1ocal region
y_o' = vapor composition leaving local region
y_e = vapor composition in equilibrium with the liquid in the local region

The proposal for calculating column vapor plate efficiencies by MacFarland, Sigmund, and Van Winkle [109] correlates with the Murphree vapor plate efficiencies in percent:

$$E_M = \frac{y_n - y_{n+1}}{y^* - y_{n+1}} \tag{19.262}$$

where

y_n = average light key mol fraction of vapor leaving plate n
y_{n+1} = average light key mol fraction of vapor entering plate n
y_e = light key mol fraction of vapor in perfect equilibrium with liquid leaving plate n

Data from bubble cap and perforated tray columns for the Murphree vapor plate efficiencies are correlated [109]:

$$E_M = 7.0(N_{DG})^{0.14}(N_{SC})^{0.25}(N_{Re})^{0.08} \tag{19.263}$$

or

$$E_M = 6.8(N_{Re}N_{SC})^{0.10}(N_{DG}N_{SC})^{0.115} \tag{19.264}$$

Referenced to 806 data points for binary systems, Equation 19.263 gives absolute deviation of 13.2%, which is about as accurate, or perhaps more so, than other efficiency equations. Equation 19.264 uses the same data and has an absolute average deviation of 10.6%. See Example 19.26 for identification of dimensionless groups.

Example 19.26: Estimating Distillation Tray Efficiency by Equations 19.263 and 19.264 [109]

Solving the problem defined in the following table will show the equations for estimating system physical properties and their relation to the calculation of Murphree vapor plate efficiencies:

System properties*	**Acetone**	**Benzene**
Molecular weight, M, lb/lb mole	58.08	78.11
Viscosity, μlb/hr-ft	0.5082	0.8155
Parachor, [P]	162.1	207.1
API specific gravity coefficient [110]		
A	0.8726	0.9485
B	0.00053	0.00053
C	21.6	18.0
E	536.0	620.6
Operating data		
Acetone mole fraction, x_1	=	0.637
Benzene mole fraction, x_2	=	0.363
Temperature, T, °F	=	166
Superficial vapor mass velocity, G, lb/hr-sq ft	=	3,820

(*Continued*)

System properties*	**Acetone**	**Benzene**
Vapor velocity, U_v, ft/hr	=	24,096
Weir height, hw, ft	=	0.2082
Fraction free area, FA	=	0.063

*Used by permission of McFarland *et al.* [109].

Liquid densities for pure hydrocarbon are calculated [110] as a function of temperature using the following equation for specific gravity:

$$s_{gL} = A - BT - C/(E - T) \tag{19.265}$$

The liquid density is then

$$\rho_L = (62.32)\ (s_{gL})$$

For acetone,

$$\rho_{L,1} = (62.32)\ [0.8726 - 0.00053\ (166) - 21.6/(536.0 - 166)] = 45.3\ \text{lb/ft}^3$$

For benzene,

$$\rho_{L,2} = (62.32)\ [0.9485 - 0.00053\ (166) - 18.0/(620.6 - 166)] = 51.2\ \text{lb/ft}^3$$

Vapor densities are calculated from the ideal gas relation:

$$\rho_V = \frac{M P_t}{555(T + 460)} \tag{19.266}$$

where total pressure P_t is given in millimeters of mercury.

Mixture densities of the binary mixtures require knowledge of volume fraction for each component. The component molar volume is

$$V_i = M_i/\rho_i \tag{19.267}$$

For acetone and benzene, respectively:

$$V_{L,1} = (58.08)/45.3 = 1.282\ \text{ft}^3/\text{lb mole}$$

$$V_{L,2} = (78.11)/1.51.2 = 1.526\ \text{ft}^3/\text{lb mole}$$

For the liquid mixture:

$$M_{L,MIX} = x_l V_{L,1} + x_2 V_{L,2} = (0.637)\ (1.282) + (0.363)\ (1.526) = 1.371\ \text{ft}^3/\text{lb mole}$$

Then the volume fraction of a component is calculated assuming an ideal mixture.

$$\upsilon_i = V_i/V_{mix} \tag{19.268}$$

For acetone and benzene, respectively:

$$\upsilon_{L,1} = 0.817/1.371 = 0.596$$

$$\upsilon_{L,2} = 0.554/1.371 = 0.404$$

The liquid density of the binary mixture is then

$$\begin{aligned}\rho_{L,MIX} &= \upsilon_{L,1}\rho_{L,1} + \upsilon_{L,2}\rho_{L,2} \\ &= (0.596)\,(45.3) + (0.404)\,(51.2) \\ &= 47.6 \text{ lb/ft}^3\end{aligned} \tag{19.269}$$

The vapor density can be found in an analogous manner.

$$\rho_{V,MIX} = \upsilon_{V,1}\rho_{L,1} + \upsilon_{V,2}\rho_{L,2} \tag{19.270}$$

However, the example problem does not require a calculation for vapor density. Instead, the superficial vapor mass velocity G can be substituted into Equation (19.278) because

$$G = U_V\rho_V \tag{19.271}$$

Liquid viscosity of the binary mixture, when not reported with the experimental efficiency results, is estimated using

$$\mu_{L,MIX} = (\Sigma x_i \mu_i^{1/3})^3 \tag{19.272}$$

The pure component viscosities are given in the literature [110, 111] as a function of temperature.

For the example,

$$\begin{aligned}\mu_{L,MIX} &= [(0.637)\,(0.5082)^{1/3} + (0.363)\,(0.8155)^{1/3}]^3 \\ &= 0.609 \text{ lb/hr-ft}\end{aligned}$$

Liquid surface tension is calculated using the Sugden Parachor method [112]. Neglecting vapor density, surface tension for the liquid mixture is

$$\sigma_{L,MIX} = \left[\left(\rho_{mix}/M_{mix}\right)[P]_{mix}\right]^4 \tag{19.273}$$

where σ is in dynes/cm, ρ is in gm/cm^3, and the parachor

$$[P]_{mix} = \Sigma x_i [P]_i \tag{19.274}$$

Values of the parachor are given in the literature [110].

Then the example gives:

$$
\begin{aligned}
M_{mix} &= (0.637)\,(58.08) + (0.363)\,(78.11)\\
&= 65.35 \text{ lb/lb mole}
\end{aligned}
$$

$$
\begin{aligned}
[P]_{min} &= (0.637)\,(162.1) + (0.363)\,(207.1)\\
&= 178.4
\end{aligned}
$$

$$\rho_{min} = 47.6/62.32 = 0.7638 \text{ gm/cm}^3$$

$$
\begin{aligned}
\sigma_{min} &= [(0.7638/65.35)\,(178.4)]^4\\
&= 18.96 \text{ dynes/cm}
\end{aligned}
$$

Diffusivity of the liquid light key component is calculated by the dilute solution equation of Wilke-Chang [113].

$$D_{LK} = (3.24 \times 10^{-8})(\psi M_{mix})^{1/2}(T + 460)/\mu_{mix}(V_{LK})^{0.6} \tag{19.275}$$

Wilke-Chang reported the recommended values for ψ as follows: water, 2.6; benzene, heptane, and ether, 1.0; methanol, 1.9; ethanol, 1.5; unassociated solvents, 1.0. The mixture parameter for the example problem is considered unity.

Then,

$$
\begin{aligned}
D_{LK} &= (3.24 \times 10^{-8})\,(65.35)^{1/2}\,(166 + 460)/(0.609)\,(1.282)^{0.6}\\
&= 2.32 \times 10^{-4} \text{ ft}^2\text{/hr}
\end{aligned}
$$

Dimensionless groups for the example problem are

$$
\begin{aligned}
N_{Dg} &= \sigma_L/\mu_L U_V\\
&= (5.417 \times 10^5)/(0.609)(2.4092 \times 104)\\
&= 37
\end{aligned} \tag{19.276}
$$

$$
\begin{aligned}
N_{Sc} &= \mu_L/\rho_L D_{LK}\\
&= (0.609)/(47.6)(2.32 \times 10^{-4})\\
&= 55
\end{aligned} \tag{19.277}
$$

$$
\begin{aligned}
N_{Re} &= h_w G/\mu_L(FA)\\
&= (0.2082)(3.82 \times 103)/(0.609)(0.063)\\
&= 2.07 \times 10^4
\end{aligned} \tag{19.278}
$$

Murphree vapor plate efficiency is calculated in two ways:

$$\begin{aligned} E_M &= 7.0\ (N_{Dg})^{0.14}\ (N_{Sc})^{0.25}\ (N_{Re})^{0.08} \\ &= 7.0\ (37)^{0.14}\ (55)^{0.25}\ (2.07 \times 10^4)^{0.08} \\ &= 7.0\ (1.66)\ (2.72)\ (2.26) = 71\% \end{aligned} \tag{19.279}$$

$$\begin{aligned} E_M &= 6.8\ (N_{Re}N_{Sc})^{0.1}\ (N_{Dg}N_{Sc})^{0.115} \\ &= 6.8\ [(2.07 \times 10^4)\ (55)]^{0.1}\ [(37)\ (55)]^{0.115} \\ &= 608\ (4.04)\ (2040) = 66\% \end{aligned} \tag{19.280}$$

In this example, Equation 19.280 gives a more conservative design basis.

where
A, B, C, E = constants in equation
D = molecular diffusion coefficient, ft^2/h
E_M = Murphree vapor plate efficiency, %
FA = fractional free area
h_w = weir height, inches
G = superficial mass vapor velocity based on the cross-sectional area of the column, lb/hr-ft^{2}t
M = molecular weight, lb/lb mole
N = dimensionless number
P = pressure, consistent units
[P] = Sugden parachor
sg = specific gravity
T = temperature, °F
U = superficial velocity, ft/h
V = molar volume, ft^3/lb mole
υ = volume fraction
x = mole fraction in the liquid
y = mole fraction in the, vapor
μ = liquid viscosity, lb/h-ft
ρ = density, lb/ft^3
σ = surface tension, dynes/cm
ψ = mixture parameter

Subscripts
i = component
L = liquid
LK = liquid light key
mix = binary mixture
n = plate number
t = total
V = vapor

Biddulph [114] emphasizes the importance of using point efficiencies rather than tray efficiencies or overall column efficiencies due to the wide fluctuations that often exist.

Kessler and Wankat [115] have examined several column performance parameters, and for O'Connell's [105] data presented in Figure 19.54, they propose equations that reportedly fit the data generally within about ±10%:

A. Distillation Trays

$$E_o = 0.54159 - 0.28531 \log_{10} \alpha\mu \tag{19.281}$$

B. Plate Absorbers (data fit ±5%)

$$E_o = 0.37237 + 0.19339 \log_{10} (HP/\mu) + 0.024816 \log_{10} (HP/\mu)^2 \tag{19.282}$$

where

E_o = overall efficiency
H' = Henry's law constant, lb mole/(atm) (ft^3)
P = pressure, atmospheres
α = relative volatility
μ = viscosity, centipoise, cP

Gerster [116] presents the results of studies on the tray efficiencies of both tray and packing contacting devices. Note that Gerster compares his work to the AIChE Manual [107].

In terms of the change in gas composition [107]:

$$E_G = E_{OG} = \frac{y - y_{n+1}}{y^* - y_{n+1}} \tag{19.283}$$

where

E_G = overall column efficiency
E_{OG} = overall point efficiency in vapor terms (see Ref. 107)
y_{n+1} = component mol fraction in the gas to the point considered
y = component mol fraction in the gas from the point considered
y^* = composition the leaving gas would have if it left the point in equilibrium with the liquid

In Table 19.27, Proctor [117] compares efficiencies of sieve and bubble cap trays (plates). He concludes that the sieve design provides a 15% improvement in plate efficiencies. To fully evaluate the actual efficiencies in any particular system, the physical properties, mechanical details of the trays, and flow rates must be considered. See also Reference 107.

Strand [118] proposes a better agreement between experimental and predicted efficiencies when recognizing a liquid by-passing factor to correct predicted values determined by the AIChE method. The results suggest that for the representative systems studied, recognition of a liquid by-passing factor for a tray can increase the agreement of the AIChE method results by as much as 5 to 10%. A vapor by-passing effect was not required to correlate the data. Because the Murphree vapor efficiencies vary considerably for various systems, the data in Reference 118 can only be a guide for other systems not studied.

This suggests that caution must be exercised when establishing a tray efficiency for any type contacting device by: (1) using actual test data if available for some similar system; (2) comparing several methods of predicting efficiency; and (3) possibly using a more conservative efficiency than calculated to avoid the possibility of ending up with a complete column with too few actual trays: a disastrous situation if not discovered prior to startup.

Sakato [119] evaluates the degree of mixing of the liquid as it flows across a tray and its effect on the tray efficiency (Figure 19.55). For plug flow, the liquid flows across the tray with no mixing, while for partial or "spot" mixing as it flows over the tray, improved tray efficiency can be expected. For a completely mixed tray liquid, the point efficiencies for a small element of the tray, E_{OG}, and tray efficiency, E_{MV}, are equal.

Table 19.27 Comparative effectiveness of sieve and bubble-cap trays/plates [178].

Plate type	Vapor throughput, Lb Mole/Hr of Dry H_2S	Over-all plate efficiency, %	
		Cold tower	Hot ower
Sieve	18,200	69 ±5	75 ± (8)*
Bubble-cap	16,200	60 ±5	69 ±5

*See the discussion of accuracy of the plate efficiency results in the text. Used by permission of the American Institute of Chemical Engineers; all rights reserved.

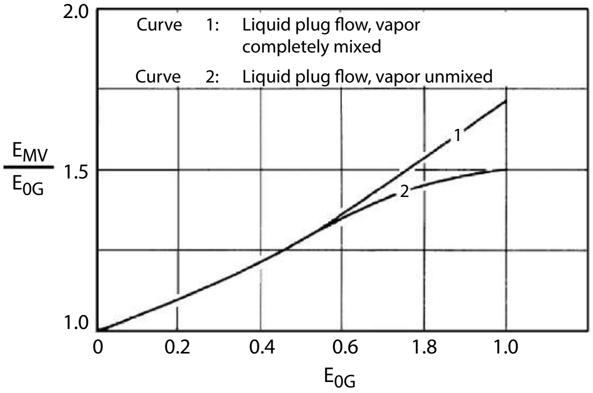


Figure 19.55 Effect of vapor mixing on tray efficiency. Reprinted by permission, Sakato, M., The American Institute of Chemical Engineers, Chem. Eng. Prog., V. 62, No. 11 (1966), p. 98, all rights reserved; reprinted by permission from Lewis, W. K., Jr., Ind. & Eng. Chem., V. 28 (1936), and by special permission from Fractionation Research, Inc.

From Figure 19.56, the effect of mixed and unmixed "pools" of liquid can be noted. For a completely mixed tray, there is no concentration gradient from inlet to outlet, and therefore, the entire tray has a uniform composition. The degree of mixing across the tray as determined by the number of discrete mixing pools on it has an effect on the relationship between E_{OG} and E_{MV} as a function of λ.

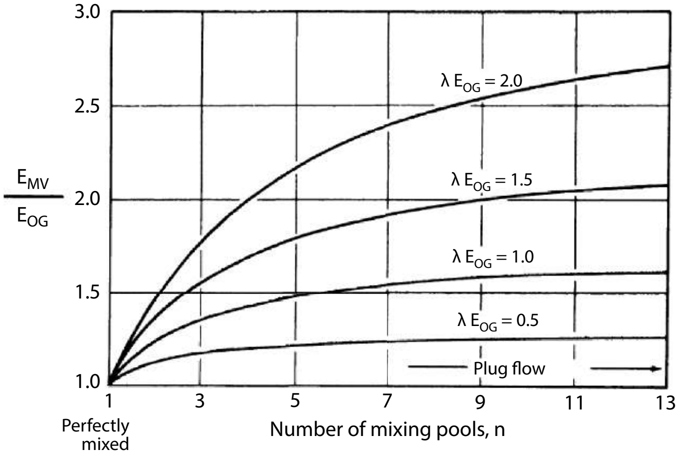


Figure 19.56 Typical effect of liquid mixing on tray efficiency. Reprinted by permission, Sakato, M., The American Institute of Chemical Engineers, Chem. Eng. Prog., V. 62, No. 11 (1966), p. 98, all rights reserved; reprinted by permission from Lewis, W. K., Jr., Ind. & Eng. Chem., V. 28 (1936), and by special permission from Fractionation Research, Inc.

where

λ = mV/L
m = slope of vapor-liquid equilibrium curve
V = vapor rate, lb mol/h
L = liquid rate, lb mol/h

Hughmark [120] has proposed empirical correlations for better fit of experimental data to transfer units and thus tray efficiency comparison with the AIChE method [107].

19.35 Steam Distillation

Live steam in steam distillation are fluids being distilled either in batch or continuous.

Ellerbe [121, 122] provides an excellent summary of the basics of steam distillation. The theory of direct steam distillation evolves around the partial pressures of the immiscible organics/petroleum/petroleum component and the presence of direct open steam in the system. The system may consist of the organic immiscible plus steam (vapor and/or liquid). Each liquid exerts its own vapor pressure independent of the other. The total pressure of the system is the sum of the individual vapor pressures of the two liquids (assuming the liquids are immiscible). An important use of this approach is to separate a volatile organic from non-organic impurities.

At constant temperature, the partial pressure for each component and the composition of the vapor phase are independent of the mols of liquid water or organic compound present; for example, for a system held at 800 mm Hg, the mixture could boil at, say, 250°F, and both liquids present would boil over together. Should one component evaporate (boil away) before the other, the system's vapor pressure would then fall to the temperature corresponding to the boiling point of the residual material.

For a system such as discussed here, the Gibb's Phase Rule [85] applies and establishes the "degrees of freedom" for control and operation of the system at equilibrium. The number of independent variables that can be defined for a system is expressed as

$$\phi + F = C + 2 \tag{19.284}$$

where

ϕ = number of phases present
F = degrees of freedom
C = number of components present

For example, for steam (saturated vapor, no liquid) distillation with one organic compound (liquid), there are two phases, two components, and hence no degrees of freedom. The degrees of freedom that could be set for the system could be: (1) temperature and (2) pressure; or (1) temperature and/or (2) concentration of the system components; or either (1) pressure or (2) concentration. In steam distillation, steam may be produced from water present, so both liquid and vapor phase water (steam) are present. For such a case, the degrees of freedom are $F = 2 + 2 - 3 = 1$.

The basic calculations involve the partial pressures of the components as discussed earlier.

For batch steam distillation: stripping [121, 122]

$$y_s = P_s/\pi \tag{19.285}$$

π = total system pressure [(also see Equation 19.3)]

$$y_{im} = p_{im}/\pi$$

The steam required for the distillation is:

$$\pi = P_s + p_{im} \tag{19.286}$$

[(also see Equation 19.2)]

$$y_s = \frac{\pi - p_{im}}{\pi}$$

The steam required per mol of immiscible liquid vaporized is:

$$\frac{y_s}{y_{im}} = \frac{\pi - p_{im}}{p_{im}} = \frac{p_s}{p_{im}} \tag{19.287}$$

$$N_s = N_{im}(\pi - P_{im)}/P_i \tag{19.288}$$

When the sum of the partial pressures of the steam and the material distilled reaches the system pressure, boiling begins and both components are vaporized in the mol ratio of their partial pressures. Upon condensation of the overhead mixture, the condensate receiver will contain two layers that can be separated by gravity.

The weight ratio of steam to the immiscible liquid in the vapor is:

$$\frac{W_s}{W_{im}} = (p_s M_s)/(p_{im} M_{im}) \tag{19.289}$$

Any non-volatile material in the mixture will be left in the still bottoms. The Hausbrand's vapor-pressure diagram [121, 122] in Figure 19.57 is a useful approach for the steam distillation calculation. This diagram was prepared for six organic compounds and the corresponding water vapor pressure as ($\pi - P_s$) for three system pressures of 760, 300, and 70 mm Hg versus temperature,

where

M = molecular weight of material
p = partial pressure, mm Hg
W = weight of material in vapor
N = number of moles

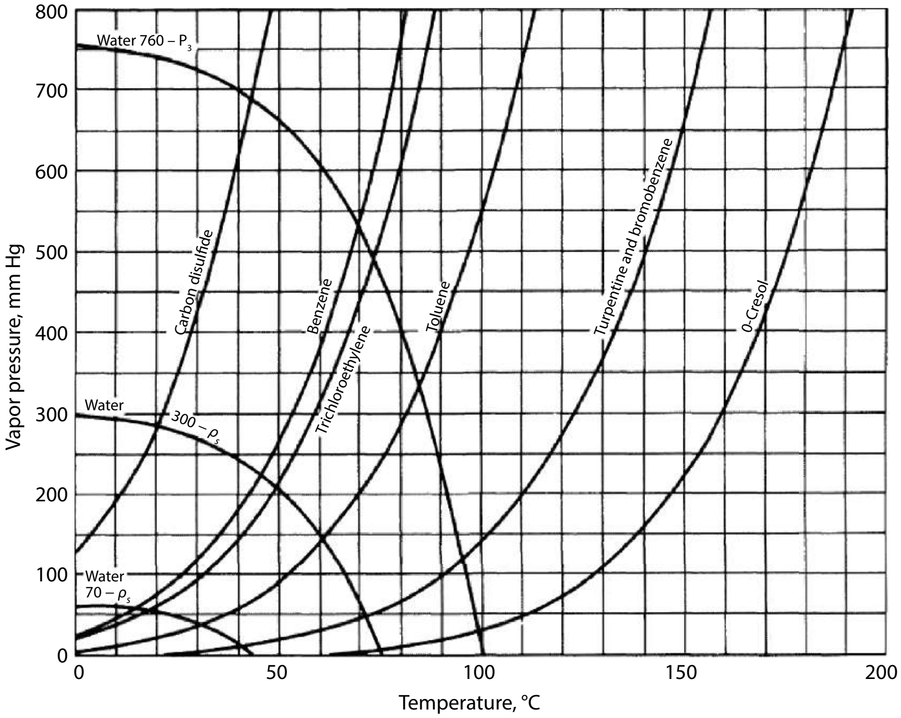


Figure 19.57 Hausbrand vapor–pressure diagram for various liquids and at three system steam pressures. A similar diagram can be constructed for other organic/hydrocarbon systems. (Used by permission, Ellerbe, R. W., Chem. Eng. March 4, p. 108, (1974)).

N_o = number of moles of non-volatile material present
y = mole fraction of material in vapor
π = system pressure, mm Hg
p_{im} = pure component vapor pressure of the immiscible liquid being distilled

Subscripts:
im = immiscible liquid
s = steam
1 = initial
2 = remaining

The water curve intersects the particular organic compound, and at that point, the temperature is that at which the steam distillation can take place, because the partial pressures are additive at this point. For example, for 300 mm Hg total pressure system, reading the intersection point for benzene and steam at 46°C gives 220 mm Hg for benzene and (300–220) or 80 mm Hg. Then the mol ratio of benzene to water vapor 220/80 = 2.75, or 2.75 parts of benzene to 1 part of water.

When the composition of the compounds in the still or bottoms changes significantly as the batch distillation progresses, an unsteady state condition will exist as for differential distillation. Batch distillation process is illustrated elsewhere [44].

When nonvolatile material in the bottoms is significant, and no liquid water exists there—that is, p_s is below saturation of the steam pressure at the still temperature—then the Raoult's Law steam efficiency is [121]: Values of E are found to range from 60% to 70% for many organics, but values of 90% to 95% are reported [121] for good sparger design for steam injection, and molecular weight of organics < 100, and 50% for many lubricating oils.

$$E = \frac{p_{im}}{p_{im}\left(\dfrac{N_{im}}{N_{im} + N_o}\right)} \tag{19.290}$$

where

E = vaporization efficiency of steam distillation

Note that for this discussion now, p_{im}, just above and in equations to follow, refers to the pure component vapor pressure of the immiscible liquid being distilled [121]. When steam is added to the still [121]:

$$\frac{p_s}{p_{im}} = \frac{\pi - p_{im}}{p_{im}} = \frac{\pi}{p_{im}} - 1 \tag{19.291}$$

and, for a constant distillation temperature, p_{im} is constant. Then for constant pressure:

$$N_S = \left(\frac{\pi}{Ep_{im}} - 1\right)(N_{im1} - N_{im2}) + \frac{\pi N_o}{Ep_{im}} \ln\left(\frac{N_{im1}}{N_{im2}}\right) \tag{19.292}$$

As the volatile material is reduced due to the batch distillation, the steam pressure rises during the progress of the operation due to the decrease of p_{im}. When the volatile material is stripped down to a low residual concentration, then p_s approaches the total system pressure, π. When the steam saturation pressure with temperature is > π, no steam condensation will occur during the operation.

When the non-volatile concentration is low to insignificant in the still feed, then N_o is small relative to N_{im}. Then p_{im} is considered constant [121]:

$$p_{im} = Ep_{im} \tag{19.293}$$

$$N_s = (p_{psp/pim}p_{im})\ (N_{iml} - N_{im2}) \tag{19.294}$$

Then, the distillate

$$\frac{W_s}{W_{im}} = \frac{(\pi - p_{im})M_s}{p_{im}M_{im}} \tag{19.295}$$

Operation of the open still with only a steam injection sparger to bring steam below the liquid level in the still may not be efficient as operating the unit with internal trays to act as a stripping column. This should be examined for each situation, as the installation of trays can be expensive, particularly if they do not aid significantly in achieving the desired separation. Generally, when no liquid water is present, the best operation of an open still (with condenser) may be at the highest working temperature achievable without adverse effects on the fluids (i.e., no polymer formation, no breakdown, etc.). Often, direct steam injection can be reduced for any operation by careful heating of the still by an internal reboiler coil, or possibly a steam jacket. The heat sensitivity of the compounds involved must be recognized.

19.35.1 Steam Distillation-Continuous Flash, Multicomponent, or Binary Mixture

This system requires direct steam injection into the still with the liquid; all the steam leaves overhead with the boiled-up vapors (no internal condensation) in a steady-state operation and system at its dew point. Steam is assumed to be immiscible with the organics. Steam distillation is usually applied in systems containing high boiling organics or heat-sensitive materials that otherwise require separation at vacuum conditions.

$$\frac{M_s}{B_{To}} = \frac{\pi}{P_b B_{To}}\left[\sum \frac{B_{io}}{\alpha_i}\right]_{i \neq s} - 1 \tag{19.296}$$

b is a more volatile reference component
$i \neq s$ = components, i, are not to include steam, s
M_s = total mols steam required
B_{To} = total mols hydrocarbons at start (not including the steam)
B_T = mols liquid in bottoms of still at time, T
B_{io} = mols of component, i, at start
α_i = relative volatility of more volatile to each of other components
P_b = vapor pressure of reference more volatile component, b
π = total system pressure, absolute
B_b = mols of component, b, used as reference for volatility, after a given time of distillation
B_o = mols of component, b, used as reference for volatility, at start of distillation

Example 19.27: Multicomponent Steam Flash Calculations

A mixture of bottoms material of composition B_{io} below has accumulated in the run-down tank. It is necessary to separate the volatile organic heavies from the tarry polymerized residue (heavy liquid). Steam is to be injected into the insulated tank containing heating coils. The system is to operate at 200 mm Hg absolute pressure and 250°F with no condensation of the steam. The organic volatile heavies contain the following:

Component	Vapor pressure @ 250°F	$\alpha = p_i/P_A$	Mols B_{io}	B_{io}/α_i
A	35 mm Hg	1.0	45	45
B	20	0.57	40	70
C	6	0.171	26	152
Total			111 mols	267

$$\frac{M_s}{B_{To}} = \frac{\pi}{P_b B_{To}}\left[\sum \frac{B_{io}}{\alpha_i}\right] - 1$$

$$\frac{M_s}{111} = \frac{200}{35(111)}[267] - 1$$

$$= 13.75 - 1 = 12.75$$

M_s = 1,415 total mols steam required for 111 mols mixture
Mole steam/mol of mixture organic volatiles = 1,415/111 = 12.8

19.35.2 Steam Distillation-Continuous Differential, Multicomponent, or Binary Mixture

The results of the differential distillation are the same as the flash distillation, although the mechanism is somewhat different. This is a batch type operation distilling differentially. All sensible and latent heats are supplied separately from the steam or by superheat in the steam. Steam acts as an inert in the vapor phase, and the quantity will vary as the distillation proceeds, while temperature and pressure are maintained.

$$\frac{M_s}{B_{T1}-B_{To}} = 1 - \frac{\pi}{P_b(B_{T1}-B_{To})}\left[\sum \frac{B_{io}}{\alpha_i}\left[\left(\frac{B_b}{B_{bo}}\right)^{\alpha i} - 1\right]\right] \tag{19.297}$$

If all the volatile materials are distilled: $\left(\frac{B_b}{B_{bo}}\right) = 0$ and $B_{T1} = 0$.

This relation is handled in a similar manner to that in flash steam separation.

If all of the material is not to be removed as overhead vapors from the still, leave a percentage of a particular compound in the bottoms, then select the particular compound as the reference material "b" for α determinations.

$$B_b = (\text{Fraction retained})(B_{bo}) \text{ and } \left(\frac{B_b}{B_{bo}}\right) = \text{Fraction retained}$$

substitute and solve for B_{T1}.

$$B_{T1} = \sum_{i\neq S} B_{io}\left(\frac{B_b}{B_{bo}}\right)^{\alpha i} \tag{19.298}$$

Knowing B_{T1}, the relation for M_s can be solved to determine mols of steam to reduce initial material to percentage of a compound in the remaining bottoms. If steam condenses, the requirement for steam increases by this amount.

19.35.3 Steam Distillation-Continuous Flash, Two Liquid Phases, Multicomponent, and Binary Mixture

Because water will be present in this system and is assumed immiscible with the other components, it will exert its own vapor pressure. This situation is similar to that with many systems where the liquid to be flashed enters below its dew point and hence requires the use of steam for heating (sensible + latent) as well as steam for the partial pressure effect.

Mole steam in vapor phase only:

$$M_s\text{ (vapor)} = P_s \sum \frac{B_{io}}{P_i} \text{ (at assumed flash temperature)}$$

where

P_s = vapor pressure of steam
P_i = vapor pressure of each component at the flash temperature

Mole steam to heat is the sum of sensible plus latent.

Total mole steam is the sum of M_s (vapor) plus heating steam.

System total pressure:

$$\pi = \frac{M_s + B_{To}}{\sum_{i\neq S} \frac{B_{io}}{P_i}}, \text{ absolute} \tag{19.299}$$

B_{To} = mole (total) volatile material at start.

19.35.4 Open Live Steam Distillation—With Fractionation Trays, Binary Mixture

Open or direct injection of steam into a distillation system at the bottom of the column or still may be used to heat the mixture as well as to reduce the effective partial pressure of the other materials. In general, if steam is used to replace a reboiler, one tray is added to replace the reboiler stage, and from 0.3 to 1.0, more trays may be needed to offset the dilution of the system with water in the lower portion. If steam is acceptable, it replaces the cost of a reboiler and any cleaning associated with this equipment. For most columns, extra trays can be purchased to offset this cost. When one of the components of the binary is water, and steam is used, the equation used for the operating stripping line (there is no rectifying section) is:

For component not including water:

$$V_s y_{i\,(m)} = L_s\, x_{i\,(m+1)} - Bx_{iB}$$

Slope of operating line $(L/V)_m = B/S$
Operating line intersects the x-axis at x_{iB}.
The step-off of trays starts at x_{iB} on the x-axis, $y = 0$.

Open steam is used for stripping of dissolved or absorbed gases from an absorption oil, with all of the steam going overhead, and the stripped oil leaving at the bottom. The absorption coefficient of the oil for the component must be known to be able to construct the equilibrium curve. The operating curve is constructed from several point material balances around the desired component, omitting the oil provided; its volatility is very low. The trays can be stepped off from a plot of y vs. x as in other binary distillations, but using only the stripping section.

Example 19.28: Continuous Steam Flash Separation Process: Separation of Non-Volatile Component from Organics Components

It is desired to separate a non-volatile material from an equimolal mixture of benzene, toluene, and xylene at 80°C. Vapor pressure data for these compounds are available from several physical property sources. The following approximate values for the specific heats and latent heats of vaporization may be used:

Benzene: $C_p = 0.419$ cal/gm-°C
$\Delta H_v = 97.47$ cal/gm

Toluene: $C_p = 0.44$ cal/gm-°C
$\Delta H_v = 86.53$ cal/gm

Xylene: $C_p = 0.40$ cal/gm-°C

$\Delta H_v = 82.87$ cal/gm

If the mixture is separated by a continuous flash process and the components are considered insoluble in water (check references) and the feed enters at the flash chamber at 20°C, calculate the mols of steam condensed, the total mols steam required per 100 mols of feed, and the total pressure. Use steam at 212°F and atmospheric pressure.

In summary, this involves a continuous flash process:

- All the feed is to be flashed.
- Steam does the heating.

- Some steam condenses.
- Water is immiscible with the materials.

Solution

Feed:

Benzene:	33.33 + mols
Toluene:	33.33 + mols
Xylene:	33.33 + mols
Total	100.00 mols feed

Because water will be present in liquid phase, it will only exert its vapor pressure. Temperature of flash = 80°C.

$$N_2 = P_s \Sigma L_i^o / P_i \qquad (19.300)$$

where

N_2 = mols steam in vapor only
P_s = vapor pressure of steam
L_i^o = mols of each component at start
P_i = vapor pressure of each component at temperature

Component	**Mols at start L_i^o**	**P_i at 80°C, mm Hg**	**L_i^o / P_i**	**Mol Wt**
Benzene	33.33	760	0.043	78
Toluene	33.33	280	0.1190	92
Xylene	33.33	120	0.277	106
			0.4397	

P_s at 80°C (176°F) = 6.868 lb/sq in. abs (from steam tables)

$$= \frac{760}{14.7}(6.868) = 354 \text{ mm Hg abs}$$

$$N_2 = P_s \sum L_i^o / P_i$$

= 354 (0.4397) = 155.7 mols steam in vapor per 100 mols of feed (volatile) material

Steam required to heat feed to 80°C:

Benzene: Sensible heat	**Btu**
(78) (33.33) [(0.419 cal/gm-°C) (1.8)](80°–20°)	= 117,800
Latent heat	
(78) (33.33) (97.46 × 1.8)	= 454,000
Toluene: Sensible heat	
(92) (33.33) [0.44 × 1.8] (80°–20°)	= 145,600
Latent heat	
(92) (33.33) (86.53 × 1.8)	= 477,000
Xylene: Sensible heat	
(106) (33.33) [0.40 x 1.8] (80°–20°)	= 153,000
Latent heat	
(106) (33.33) (82.87 × 1.8)	= 525,000
Total heat load:	= 1,872,400

lbs steam required for heat load

$$=\left[\frac{1{,}872{,}400\ \text{Btu/100mol}}{970\ \text{Btu/lb at } 212°\text{F}}\right]=1{,}932\ \text{lbs steam/100mol}$$

$$\text{Mole steam required for heat load} = \frac{1{,}932}{18}=107.2\ \text{ mole steam/100 mols volatile material}$$

$$\text{Total mole steam/100 mole volatile feed} = 155.7 + 107.2 = 262.9$$

Total pressure of system:

$$\pi=\frac{N_2+L_T^o}{\sum_{i\neq S} L_i/P_i}$$

$$L_T = \text{Total mole volatile material at start} = 100$$

$$\pi=\frac{155.7+100}{0.4397}=580\ \text{mm Hg abs}$$

Example 19.29: Open Steam Stripping of Heavy Absorber Rich Oil of Light Hydrocarbon Content (Used by Permission Following the Method of R. W. Ellerbe, Chemical Engineering [121])

A gas processing plant selectively extracts ethylene and ethane from an incoming natural gas mixture stream. These two light hydrocarbons are absorbed in a heavy gasoline type absorber “oil” and then stripped with open steam in an open tower. The system data are as follows (see Figure 19.58):

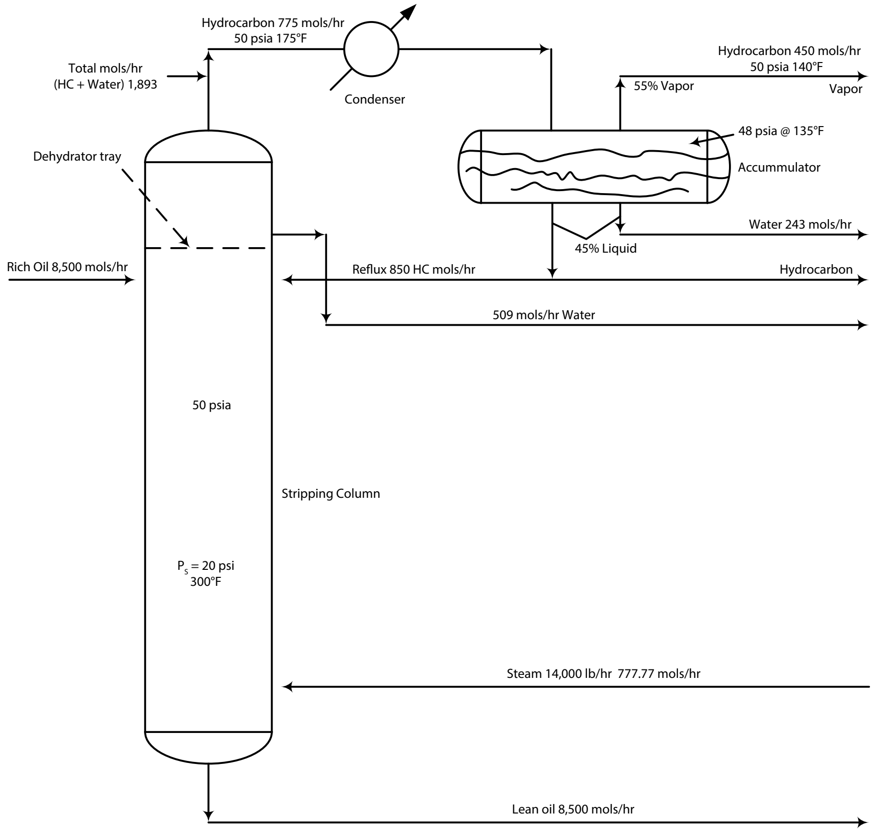


Figure 19.58 Open steam stripping light hydrocarbons from rich oil. Modified for Example 19.29 and used by permission, Ellerbe, R. W., Chem. Eng. Mar. 4, p. 108, 1974.

Rich oil rate to tower:	8,500 mol/h
Overhead product of ethylene and ethane:	775 mol/h
Overhead product from accumulator:	55% vapor and 45% liquid
Accumulator conditions:	48 psia and 135°F
Reflux hydrocarbon in top vapor @ 175°F:	850 mol/h
Steam (superheated) enter bottoms below tray:	14,000 lb/h
Water partial pressure in the mixed vapor at bottoms:	20 psi
Hydrocarbons mix partial pressure:	50 psi – 20 psi = 30 psi

Neglect pressure drops through the system.

Determine: How much water is removed from the overhead accumulator and the intermediate dehydrator or water removal tray? No water is removed from the bottoms due to the use of superheated steam.

Solution

From steam tables (saturated) at:

Top of tower, 175°F, vapor pressure water, psia	= 6.8
Mole fraction water vapor at top of tower: 6.8/48 psia	= 0.1416
Mole fraction hydrocarbon at top of tower: 1 – 0.1416	= 0.8584
Total mole mix HC vapor and water vapor at tower overhead: $\frac{775+850}{0.8584}=1{,}893.0$	
Mole of water vapor in tower overhead: 1,893 – (775 + 850)	= 268
Accumulator @ 48 psia & 135°F, water vapor pressure:	= 2.6 psia
Mole fraction water in accumulator vapor: = 2.6/48	= 0.0541
Mole fraction HC in accumulator vapor: = 1 – 0.0541	= 0.9459
Total mole vapor leaving accumulator: = $\left[\frac{(775)(0.55)}{0.8584}\right]$	= 450.6
Mole water vapor leaving accumulator: = 450.6 – (775) (0.55)	= 24.35
Mole liquid water withdrawn from accumulator: = 268 – 24.35	= 243.65
Mole liquid water collected on dehydrator tray and removed at that point up tower above	= 777.7 - 268
where reflux returns below this tray: (water vapor in tower overhead)	= 509.7 mole/h
Mole steam entering tower: = 14,000/18	= 777.7 mole/h

19.36 Distillation with Heat Balance of Component Mixture

This type of evaluation of a distillation system involves a material and heat balance around each tray. This is extremely tedious by conventional means, but is now handled with computers using simulation software (e.g., Hysys, Chemcad, Prosim, UniSim, Aspen). However, the volume of calculations is large and requires a relatively long time. Only those special systems that defy a reasonable and apparently economical solution by other approaches are even considered for this type of approach. The detailed method involves trial and error assumptions of both material balance and heat balance.

19.36.1 Unequal Molal Overflow

This is another way of expressing that the heat load from tray to tray is varying in the column to such an extent as to make the usual simplifying assumption of equal molal overflow invalid. The relations to follow do not include heats of mixing. In general, they apply to most hydrocarbon systems.

1. Equation of operating line in rectifying section, light component [85].

$$L_{n+1} = V_n - D$$

$$y_n = \left(\frac{M_D - H_n}{M_D - h_{n+1}}\right) x_{n+1} + \left(\frac{H_n - h_{n+1}}{M_D - h_{n+1}}\right) x_D$$

$$\frac{L_{n+1}}{V_n} = \frac{M_D - H_n}{M_D - h_{n+1}} = 1 - \frac{H_n - h_{n+1}}{M_D - h_{n+1}} \qquad (19.301)$$

2. Equation of operating line in stripping section, light component

$$
\begin{aligned}
L_{m+1} &= V_m + B \\
y_m &= \left(\frac{M_B - H_m}{M_B - h_{m+1}}\right)x_{m+1} + \left(\frac{H_m - h_{m+1}}{M_B - h_{m+1}}\right)x_B \\
\frac{L_{m+1}}{V_m} &= 1 - \frac{H_m - h_{m+1}}{M_B - h_{m+1}} \\
\frac{D}{V_m} &= \frac{H_m - h_{m+1}}{M_m - h_{n+1}}
\end{aligned}
\tag{19.302}
$$

where

$M_B = h_B - Q_B/B$

$M_D = Q_c/D + h_D$

$M_B = h_w - \frac{Q_s}{W}$

H_n = total molal enthalpy of vapor at conditions of plate n, $H_n = \Sigma H_{ni}\,(y_{ni})$

h_n = total molal enthalpy of liquid at conditions of plate n, $h_n = \Sigma h_{ni}\,(x_{ni})$

s = lb (or mole) steam per lb (or mol) bottoms

H_m = total molal enthalpy of vapor at plate m (below feed)

N = mole residue or bottoms per unit time

Q_B = heat added in still or bottoms

19.36.2 Ponchon–Savarit Method-Binary Mixtures

For a binary fractionating tower, illustrated in Figure 19.59, the conditions of the feed, F, x_F, H_F are known, and the composition of the distillate product x_D and the composition of the bottoms product x_B are specified. If the column is operated at a constant reflux ratio, the number of equilibrium stages required to achieve a desired separation can be estimated. The simplest method of achieving this task is based upon an assumption of constant molal overflow (CMO). This only occurs if:

1. The molal heats of vaporization are negligible.
2. Sensible heat distributions are negligible.
3. Heats of mixing are negligible.

With these criteria, the internal energy balances around each plate in the tower are straightforward, as the molal flow rate of liquid and vapor is constant throughout each section. The McCabe–Thiele diagram can be employed to calculate the number of equilibrium stages. However, the above criteria are not met for numerous systems. The use of McCabe–Thiele analysis in these circumstances may lead to under specification of distillation equipment.

Where constant molal overflow is not applicable, enthalpy-composition diagrams can be combined with vapor-liquid equilibrium data and the graphical Ponchon-Savarit method employed. A Ponchon-Savarit analysis does not assume the liquid and vapor flow rates are constant in each section of the tower. The only assumption of this approach is that there are no significant heat losses through the column walls.

This graphical method allows solution of many distillation problems, which would require considerable work if attempted by rigorous methods. Robinson and Gilliland have published technical and descriptive details substantiating this method [85, 123]. Figure 19.60 presents a summary of the use of this method and appropriate interpretations. Scheiman [124] uses the Ponchon–Savarit diagrams to determine minimum reflux by heat balances. Campagne [125, 126] suggests a detailed technique for using the Ponchon–Savarit method with a computer simulation, which leads to designs not previously possible.

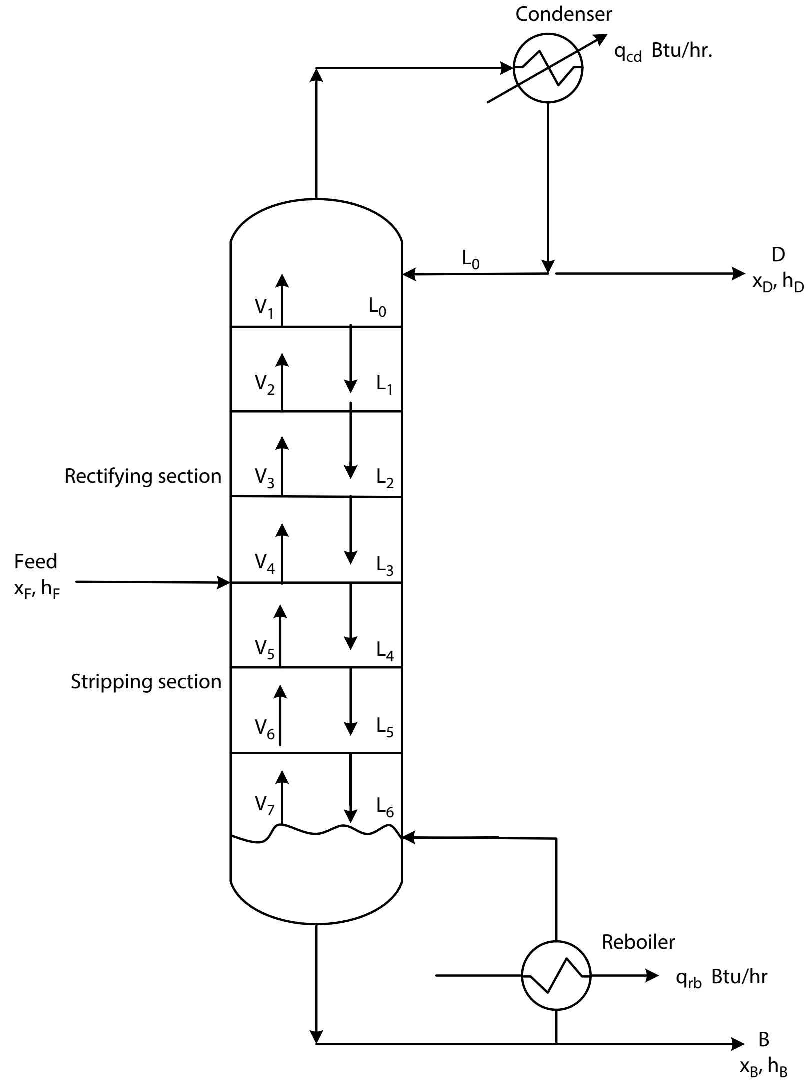


Figure 19.59 Fractionating tower with reboiler and condenser units.

The basic method allows the non-ideal heat effects of the system to be considered as they affect the plate-to-plate performance. The systems, as represented in the diagrams, are usually at constant pressure, but this is not necessarily the case in reality. The equilibrium tie lines connect points fixed by the x–y values to corresponding saturated liquid and saturated vapor conditions at a constant temperature, such as "a" °F or "b" °F. The mol fractions are obtained from the usual x–y diagram for the system, and the enthalpy values are relative to a fixed datum point for the available heat data of the particular components. For such systems as ammonia–water, methanol–water, and ethanol–water, the data are readily available (see Figures 19.61a, 19.62, and 19.63). The saturated liquid line represents the enthalpies of liquid mixtures at the various compositions, all at a constant pressure. This is the bubble point curve. The dew point curve is produced by plotting the enthalpies of the various vapor mixtures at the saturation temperature at a constant pressure. Figure 19.61b shows the corresponding vapor-liquid equilibrium diagram for ammonia–water system at 1.01325 kPa, and Figure 19.64 illustrates the combined enthalpy-concentration diagrams for ethanol–water system at 1.01325 kPa.

An effort has been made to present the basic understanding of the method as it applies to systems involving unequal molal overflow, open steam distillation, and single flash vaporization in Figures 19.60 and 19.65.

To obtain high or even necessary levels of accuracy for some design conditions, the end portions of the graphical representation may require enlargement from the usual size for graphical plotting. In most cases, a size of 11 × 17

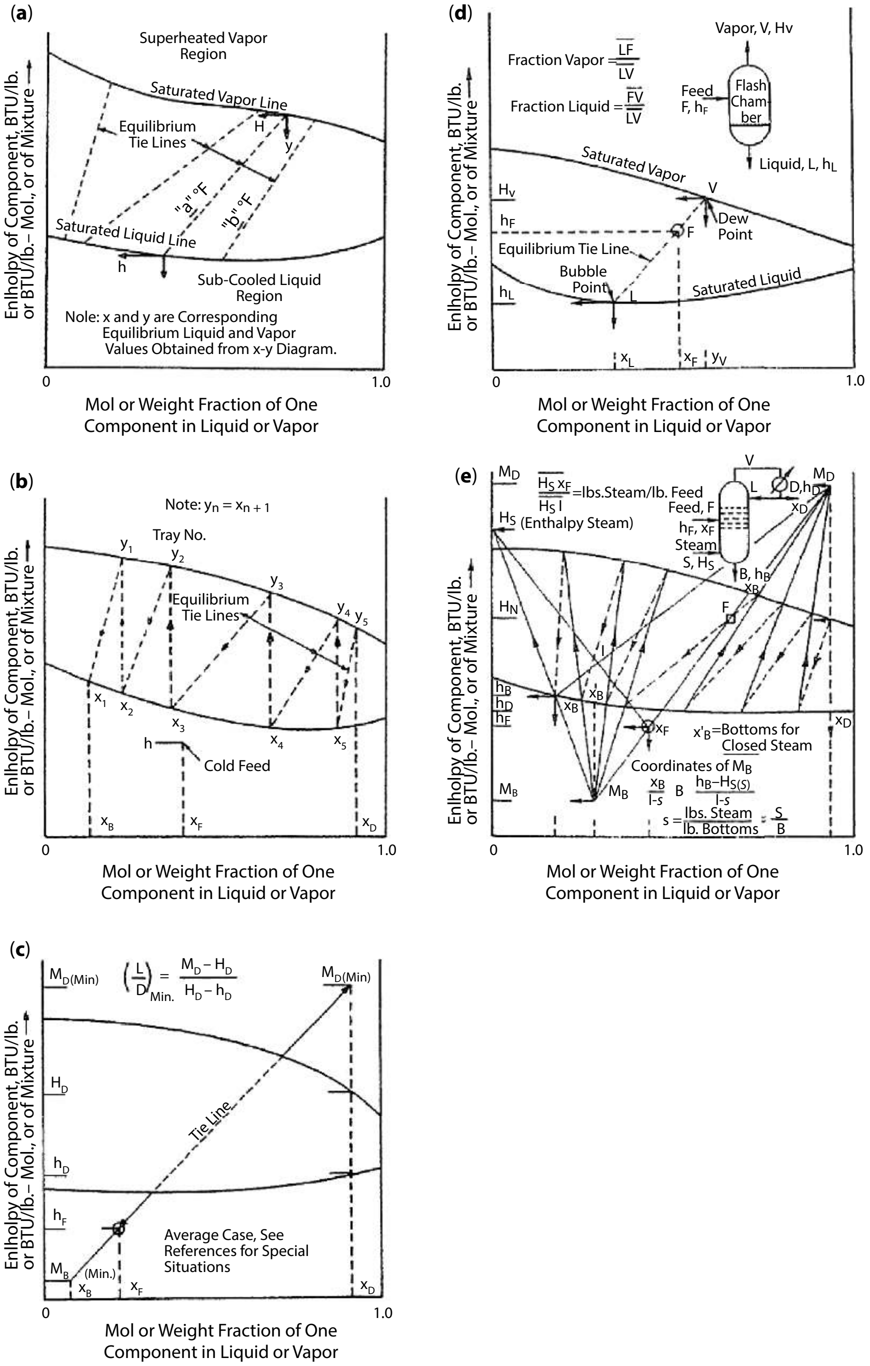


Figure 19.60(a-e) Performance analysis of unequal mol overflow to binary system using Ponchon–Savarit method.

inches is suggested. In general, the nonequimolal overflow (NEMO) model of the column is more vigorous and should be used in those columns where there are:

1. Large temperature differences from the top of the column to the bottom (i.e., where sensible heat effects are important).
2. Large heats of mixing.
3. Significant differences in the molal heats of vaporization of the components.

The Equations

Consider the energy balance around the column, assuming no heat loss through the column wall:

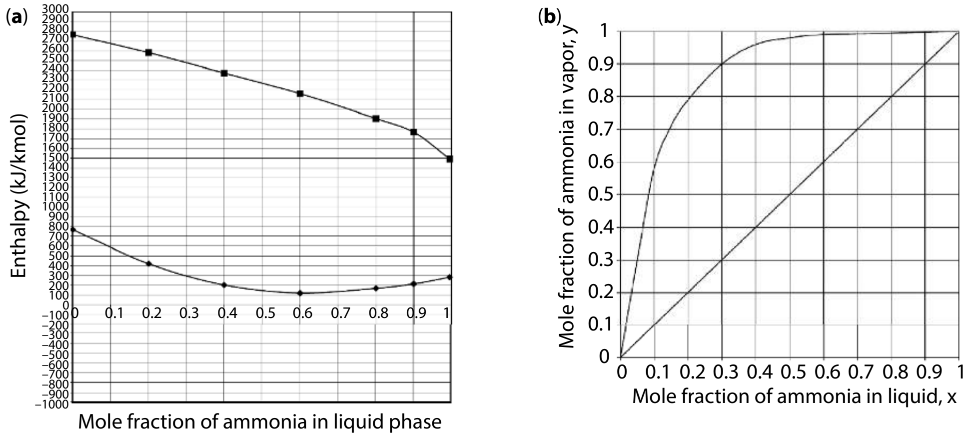


Figure 19.61 (a) Enthalpy vs composition of ammonia–water system. (b) VLE data of ammonia–water system at 1atm.

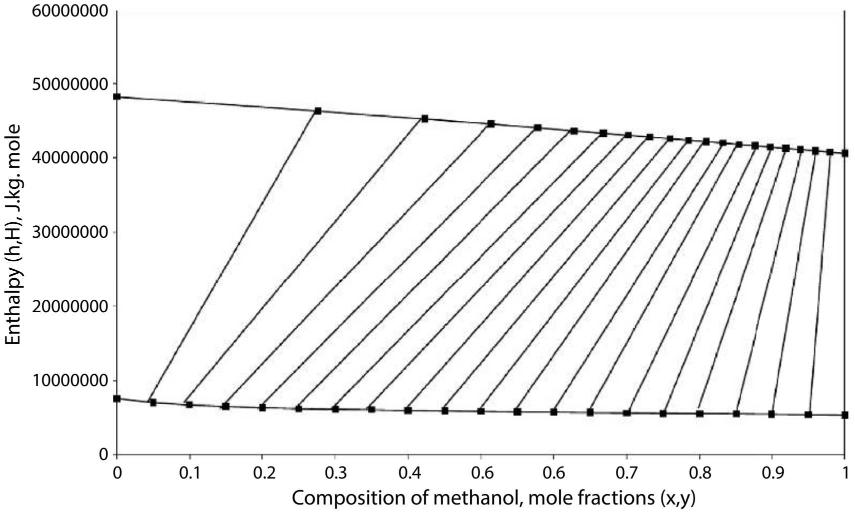


Figure 19.62 Enthalpy composition diagram of methanol–water mixture at a pressure of 101.235 kPa.

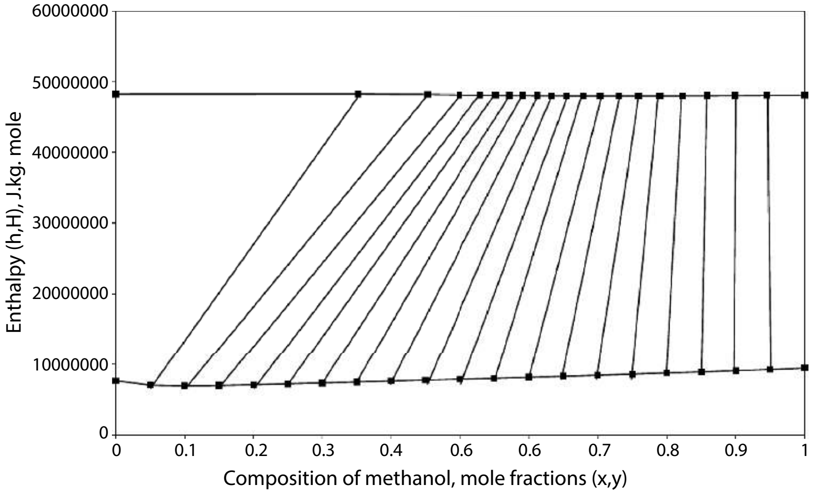


Figure 19.63 Enthalpy composition diagram for ethanol–water mixture at a pressure of 101.235 kPa.

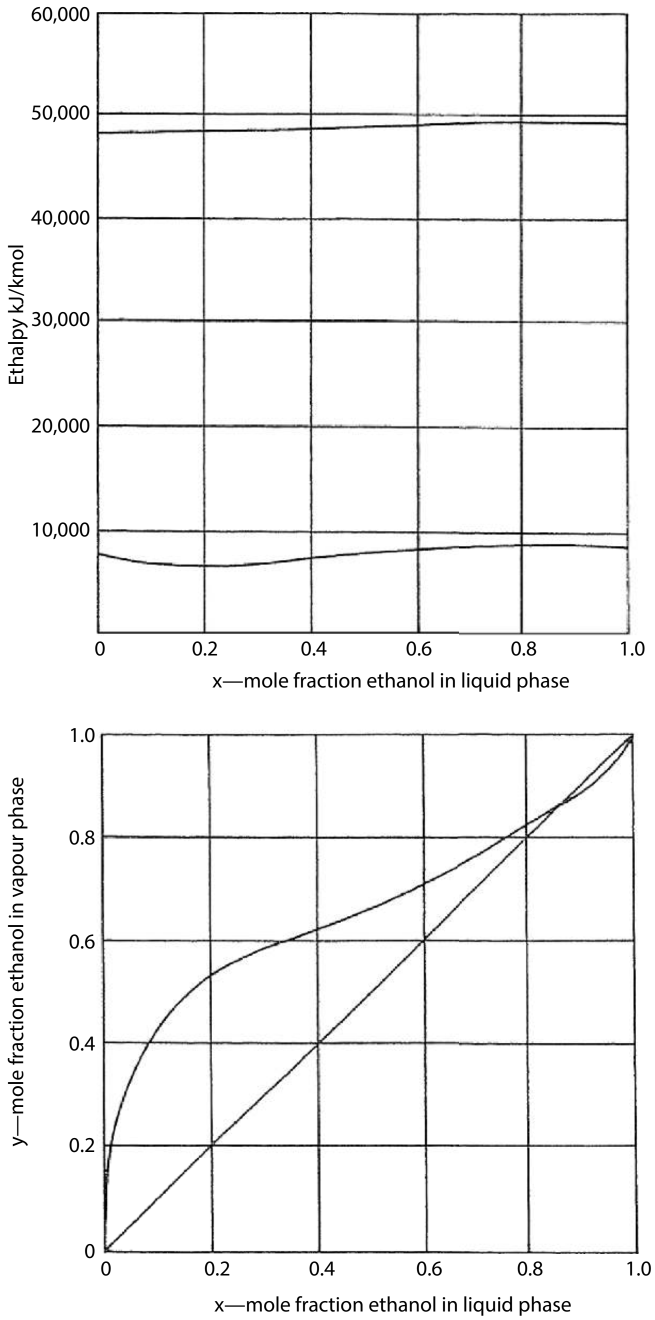


Figure 19.64 Combined enthalpy-concentration diagrams for ethanol–water system at 1.0 bar.

$$Fh_f + q_{rb} = Dh_D + Bh_B + q_{cd} \tag{19.303}$$

where q_{rb} and q_{cd} are the duties of the reboiler and condenser, respectively, Btu/h. Since the column reflux ratio R is set, R is defined by

$$R = L/D \tag{19.304}$$

or

$$L = RD \tag{19.305}$$

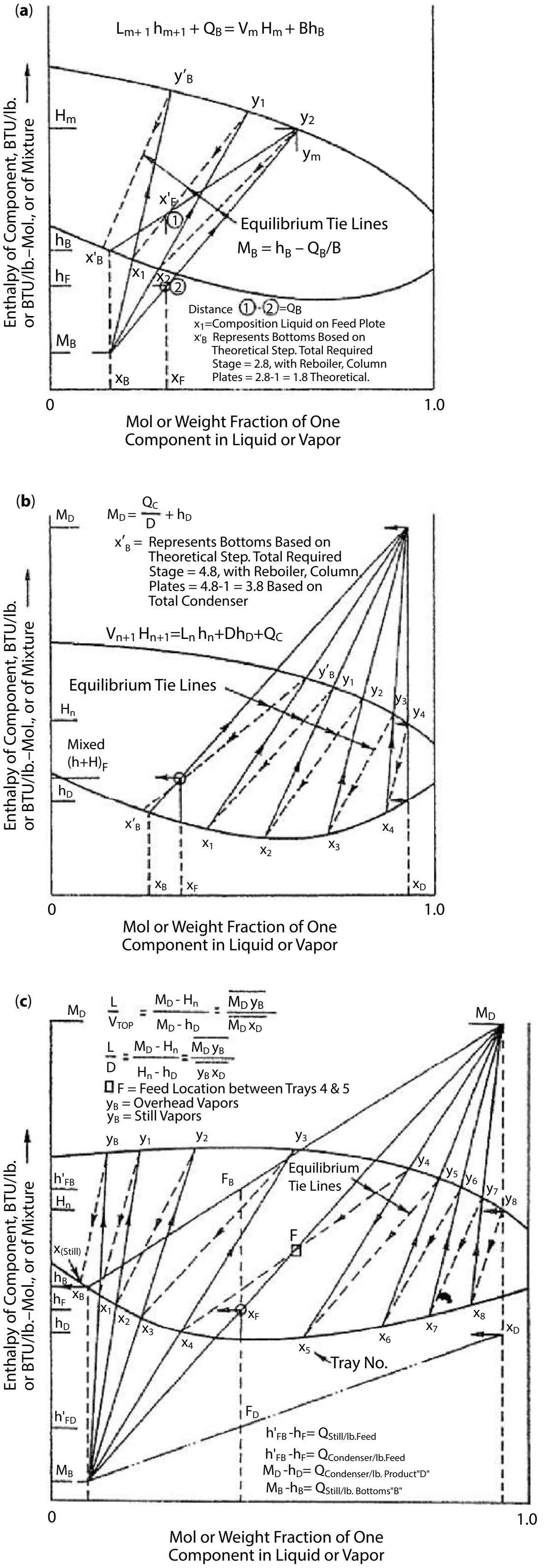


Figure 19.65(a-c) Graphical solution of unequal molal overflow, binary systems.

where L = the molal flow rate of reflux returned to the column, the condenser duty is

$$q_{cd} = (L + D)\lambda \tag{19.306}$$

Substituting Equation (19.305) into Equation (19.306) gives:

$$\begin{aligned} q_{cd} &= (RD + D)\lambda \\ &= D(R + 1)\lambda \end{aligned} \tag{19.307}$$

where
 λ is the latent heat of vaporization of the distillate product.

The distillate product flow rate D and the bottoms product flow rate B can be determined from an overall mass balance and a mass balance on the more volatile component as follows:

$$F = D + B \tag{19.308}$$

$$Fx_F = Dx_D + Bx_B \tag{19.309}$$

Since the feed stream and the compositions of distillate and bottoms product have been specified, the two unknown quantities D and B in Equations (19.308) and (19.309) can be found.

Combining Equations (19.308) and (19.309) gives:

$$(D + B)x_F = Dx_D + Bx_B \tag{19.310}$$

Rearranging Equation (19.310) gives

$$B(x_F - x_B) = D(x_D - x_F) \tag{19.311}$$

From the heat balance of Equation (19.303),

$$Fh_F + q_{rb} = Dh_D + Bh_B + q_{cd} \tag{19.303}$$

Rearranging Equation (19.303) gives

$$Fh_F = \left(h_B - \frac{q_{rb}}{B}\right)B + \left(h_D + \frac{q_{cd}}{D}\right)D \tag{19.312}$$

Rearranging Equations (19.311) in the ratio of B/D gives

$$\frac{B}{D} = \frac{x_D - x_F}{x_F - x_B} \tag{19.313}$$

Substituting Equation (19.308) into Equation (19.312) gives

$$(D + B)h_F = \left(h_B - \frac{q_{rb}}{B}\right)B + \left(h_D + \frac{q_{cd}}{D}\right)D \tag{19.314}$$

or

$$\left(1 + \frac{B}{D}\right)h_F = \left(h_B - \frac{q_{rb}}{B}\right)\frac{B}{D} + \left(h_D + \frac{q_{cd}}{D}\right) \tag{19.315}$$

Substituting Equation (19.313) into Equation (19.315) gives

$$\left[1+\left(\frac{x_D - x_F}{x_F - x_B}\right)\right]h_F = \left(h_B - \frac{q_{rb}}{B}\right)\left(\frac{x_D - x_F}{x_F - x_B}\right) + \left(h_D + \frac{q_{cd}}{D}\right) \tag{19.316}$$

$$h_F + \left(\frac{x_D - x_F}{x_F - x_B}\right)h_F = \left(h_B - \frac{q_{rb}}{B}\right)\left(\frac{x_D - x_F}{x_F - x_B}\right) + \left(h_D + \frac{q_{cd}}{D}\right) \tag{19.317}$$

Rearranging Equation (19.317) gives

$$\left(\frac{x_D - x_F}{x_F - x_B}\right)\left[h_F - \left(h_B - \frac{q_{rb}}{B}\right)\right] = \left(h_D + \frac{q_{cd}}{D}\right) - h_F \tag{19.318}$$

and

$$\frac{x_F - x_B}{\left[h_F - \left(h_B - \frac{q_{rb}}{B}\right)\right]} = \frac{x_D - x_F}{\left(h_D + \frac{q_{cd}}{D}\right) - h_F} \tag{19.319}$$

Equation (19.319) indicates that the points $x_B, (h_B - q_{rb}/B)$; h_F, x_F; and $x_D, (h_D + q_{cd}/D)$ must all lie on a single straight line on an enthalpy–composition diagram. Since we know the condenser duty, q_{cd}, the point x_D, $(h_D + q_{cd}/D)$ can be positioned on the enthalpy composition diagram as point d in Figure 19.66a-c. Place the feed point f at h_F, x_F and extend a line through points d and f to a vertical line placed at $x = x_B$ to point b. This point must have the coordinates $x_B, (h_B - q_{rb}/B)$.

Note: The feed is taken as saturated liquid.

Proceed to calculate the number of equilibrium stages in the rectifying section as follows: First, draw a vertical line at $x = x_D$ extending from point d to the saturated line at point l_0. Both the distillate product and reflux are saturated liquid with composition x_D and hence must be represented by point l_0 (Figure 19.66b).

The mass balance, the component balance, and the energy balance around the condenser are

$$V_1 = L_1 + D \tag{19.320}$$

$$V_1 y_1 = L_1\, x_1 + Dx_D \tag{19.321}$$

$$V_1\, h_v = L_1\, h_L + D\left(h_D + \frac{q_{cd}}{D}\right) \tag{19.322}$$

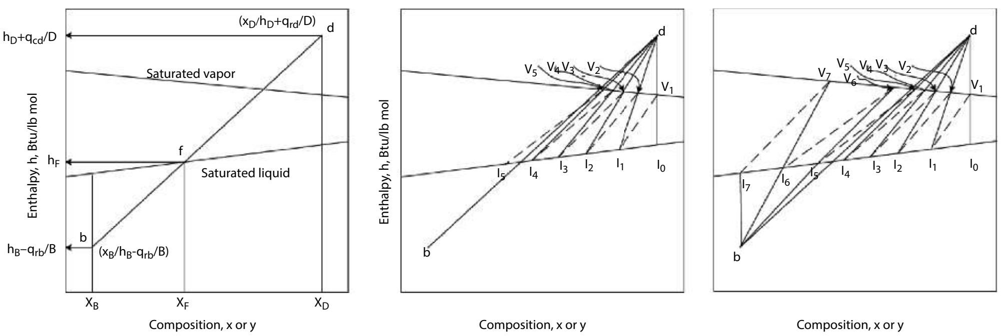


Figure 19.66(a–c) Ponchon analysis of a fractionating column.

Around the top plate of the column and the condenser, an energy balance yields

$$V_2\,h_{v2} = L_1\,h_{L1} + D\left(h_D + \frac{q_{cd}}{D}\right) \tag{19.323}$$

where

hv = enthalpy of saturated vapor
h_L = enthalpy of saturated liquid

The total and component mass balances are

$$V_2 = L_1 + D \tag{19.324}$$

$$V_2 y_2 = L_1\,x_1 + Dx_D \tag{19.325}$$

Substituting V_2 in Equation (19.324) into Equations (19.323) and (19.325), respectively, gives

$$(L_1 + D)y_2 = L_1 x_1 + Dx_D \tag{19.326}$$

$$(L_1 + D)h_{v2} = L_1 h_{L1} + D\left(h_D + \frac{q_{cd}}{D}\right) \tag{19.327}$$

Rearranging Equation (19.327) gives

$$\left(\frac{L_1}{D} + 1\right)h_{v2} = \frac{L_1}{D}h_{L1} + \left(h_D + \frac{q_{cd}}{D}\right) \tag{19.328}$$

Rearranging Equation (19.326) gives

$$L_1 y_2 - L_1 x_1 = Dx_D - Dy_2 \tag{19.329}$$

or

$$L_1(y_2 - x_1) = D(x_D - y_2) \tag{19.330}$$

$$\frac{L_1}{D} = \frac{x_D - y_2}{y_2 - x_1} \tag{19.331}$$

Substituting Equation (19.331) into the energy balance Equation (19.328) gives

$$\left(\frac{x_D - y_2}{y_2 - x_1} + 1\right)h_{v2} = \left(\frac{x_D - y_2}{y_2 - x_1}\right)h_{L1} + \left(h_D + \frac{q_{cd}}{D}\right) \tag{19.332}$$

or

$$(h_{v2} - h_{L1})\left(\frac{x_D - y_2}{y_2 - x_1}\right) = \left(h_D + \frac{q_{cd}}{D}\right) - h_{v2} \tag{19.333}$$

$$\frac{L_1}{D}=\frac{x_D-y_2}{\left(h_D+\frac{q_{cd}}{D}-h_{v2}\right)}=\frac{y_2-x_1}{h_{v2}-h_{L1}} \tag{19.334}$$

The familiar form of Equation 19.334 indicates that points d and l_1 in Figure 19.66b must be on the same straight line as points y_2, h_{v2}. Since the location of points d and l_1 is known, a line can be drawn connecting these points. Stream V_2 is saturated vapor, and the intersection of line d–l_1 with the saturated vapor line at point V_2 must represent the enthalpy and the composition of stream V_2 in Figure 19.66b. The process is repeated to calculate the enthalpy and composition of streams L_2 and V_3. A tie line is drawn from V_2 to the saturated liquid line. Point l_2 thus determined specifies stream L_2. From point l_2, a line is constructed to point d. By reasoning analogous to that following Equation 19.334, stream V_3 is represented by the intersection of line l_2d with the saturated vapor line at point V_3. The process is continued for each plate in the rectifying section of the column and the number of plates determined by the number of tie lines that were constructed.

When a tie line crosses line ***bfd*** (Figure 19.66), the feed point is determined. A new set of material and energy balances is then used to find the number of plates in the stripping section. This corresponds to using different operating lines in the stripping and rectifying sections of a column in a McCabe–Thiele analysis. It is not essential that the feed plate is chosen whose tie line crosses line ***bfd*** as the feed plate. However, with this criterion for locating the feed plate, the maximum separation per plate is achieved.

Stripping Section (see Figure 19.59)

The number of plates in the stripping section of the column (i.e., plates below the feed point) is determined in the same manner as in the rectifying section, and the procedure is as follows:

An energy balance is written around the second plate in the stripping section and the reboiler:

$$h_4 h_{L4}=V_5\,h_{v5}+B\left(h_B-\frac{q_{rb}}{B}\right) \tag{19.335}$$

The mass and component balances are

$$L_4 = V_5 + B \tag{19.336}$$

$$L_4\,x_4 = V_5\,y_5 + B\,x_B \tag{19.337}$$

Combining Equations (19.336) and (19.337) gives

$$(V_5 + B)x_4 = V_5 y_5 + Bx_B \tag{19.338}$$

Rearranging Equation (19.338) gives

$$V_5(y_5 - x_4) = B(x_5 - x_B) \tag{19.339}$$

Therefore,

$$\frac{B}{V_5}=\frac{y_5-x_4}{x_5-x_B} \tag{19.340}$$

From the heat balance,

$$(V_5 + B)h_{L4} = V_5 h_{v5} + B\left(h_B - \frac{q_{rb}}{B}\right) \tag{19.341}$$

or

$$\left(1 + \frac{B}{V_5}\right)h_{L4} = h_{v5} + \frac{B}{V_5}\left(h_B - \frac{q_{rb}}{B}\right) \tag{19.342}$$

Substituting Equation (19.330) into Equation (19.342) gives

$$\left(1 + \frac{y_5 - x_4}{x_4 - x_B}\right)h_{L4} = h_{v5} + \left(\frac{y_5 - x_4}{x_4 - x_B}\right)\left(h_B - \frac{q_{rb}}{B}\right) \tag{19.343}$$

Rearranging Equation (19.343) gives

$$\left(\frac{y_5 - x_4}{x_4 - x_B}\right)\left[h_{L4} - \left(h_B - \frac{q_{rb}}{B}\right)\right] = h_{v5} - h_{L4} \tag{19.344}$$

or

$$\frac{y_5 - x_4}{h_{v5} - h_{L4}} = \frac{x_4 - x_B}{\left[h_{L4} - \left(h_B - \frac{q_{rb}}{B}\right)\right]} \tag{19.345}$$

Rearranging Equation (19.345) for the stripping section of the column finally gives

$$\frac{B}{V_5} = \frac{y_5 - x_4}{x_4 - x_B} = \frac{h_{v5} - h_{L4}}{\left[h_{L4} - \left(h_B - \frac{q_{rb}}{B}\right)\right]} \tag{19.346}$$

In the stripping section, from Equation (19.346), points b and l_4 must lie on a straight line with points h_{v5} and y_5. Since the location of points b and l_4 is known, a line is extended through these points to the saturated vapor line to determine points V_5 and the enthalpy and composition of a stream V_5 as shown in Figure 19.66c.

Streams V_5 and l_5 are in equilibrium, and a tie line from point V_5 to the saturated liquid line set point l_5 and the conditions of stream l_5. This process is continued by drawing a line through points b and l_5 to find point V_6 on the saturated vapor line. By repeating this process until a tie line intersects the saturated liquid line at or below x_B, we complete the Ponchon–Savarit analysis. The number of equilibrium stages in the fractionating column is determined by counting the number of tie lines that were constructed. In Figure 19.66c, a total of seven equilibrium stages, three plates in the rectifying section, one feed plate, and three plates in the stripping section, corresponds to the column reboiler. Points d and b in Figures 19.66a-c are referred to as difference points, and lines drawn to d and b are referred to as difference lines.

Minimum Reflux

The condition of minimum reflux is when any equilibrium tie line coincides with any operating line, as shown in Figure 19.67. Usually this coincidence takes place on the operating line through the feed condition, and the location of d for the condition of minimum reflux is found by extending the equilibrium tie line through the feed condition to intersect the composition through x_D or y_D.

Total Reflux

The condition of total reflux is shown in Figure 19.68. In this instance, the operating lines are parallel and vertical, converging at infinity. This results from the condition of V = L since D = 0, B = 0, $Q_D = q_{cd}/D = \infty$, and $Q_B = q_{rb}/B = \infty$. Thus, the enthalpy coordinates are $h_D - Q_D = h_D - (-\infty)$ and $h_B - Q_B = h_B - (\infty)$.

The minimum number of plates is encountered at the condition of total reflux.

Side Streams and Multiple Feeds

The Ponchon–Savarit method is very versatile and can be applied with columns having multiple feeds or side stream withdrawals. There are situations where feeds of different composition are involved, and it is possible to introduce each at its proper location in the column, and thereby reduce the total number of plates by one or two compared when the feeds are premixed. Alternatively, process requirements may require two or more distillate products of different composition, and these may be removed from the column as distillate and side streams. The Ponchon–Savarit method can be employed in these cases by writing the difference – point equations for each section of the column between streams added or removed. The number of difference points will always be equal to the number of streams N (including the distillation and bottoms) minus 1 as shown in Figures 19.69 and 19.70.

Columns with side streams may be handled in much the same way as a single feed stream. Here, three difference points are needed to represent the three sections of the column in Figure 19.70. Points d_1 and f′ are located as shown, with f′ representing a hypothetical stream in which a side stream of saturated liquid has been physically subtracted from the saturated liquid feed. Using the level arm principle, point f′ must lie on a line extended through points s and f such that F/S = f′s/f′f. Similarly, in the case of multiple feeds, the difference point b is found by the intersection

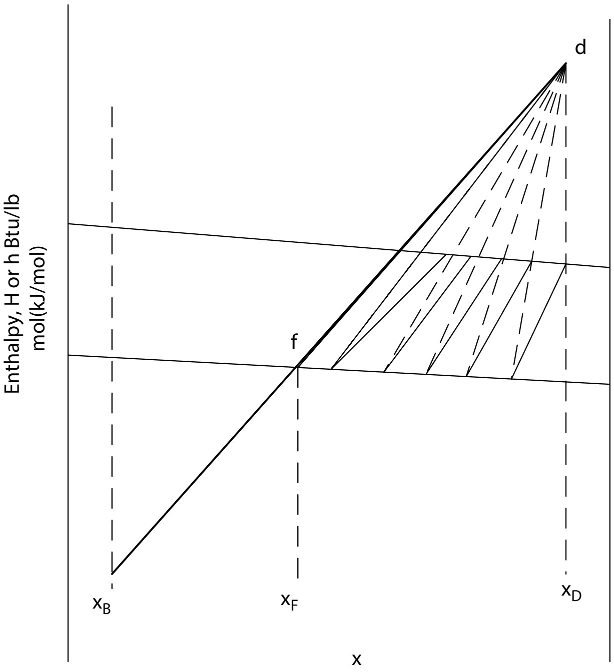


Figure 19.67 Minimum reflux ratio on an enthalpy–concentration diagram.

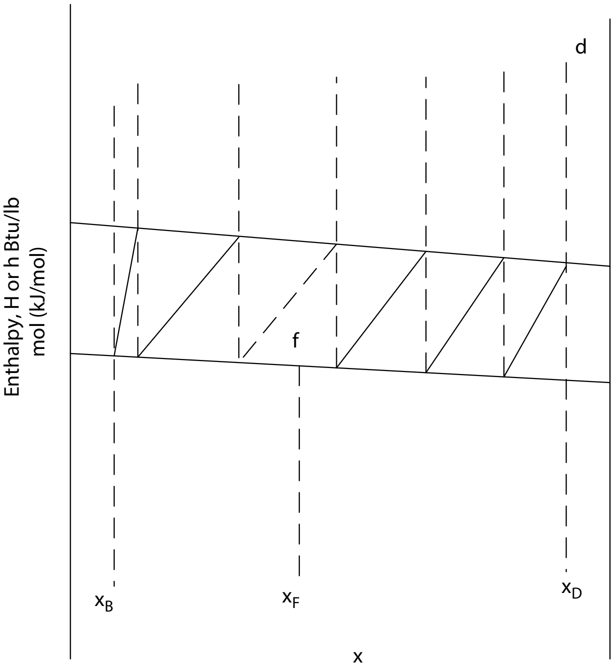


Figure 19.68 Total reflux construction on an enthalpy–concentration diagram.

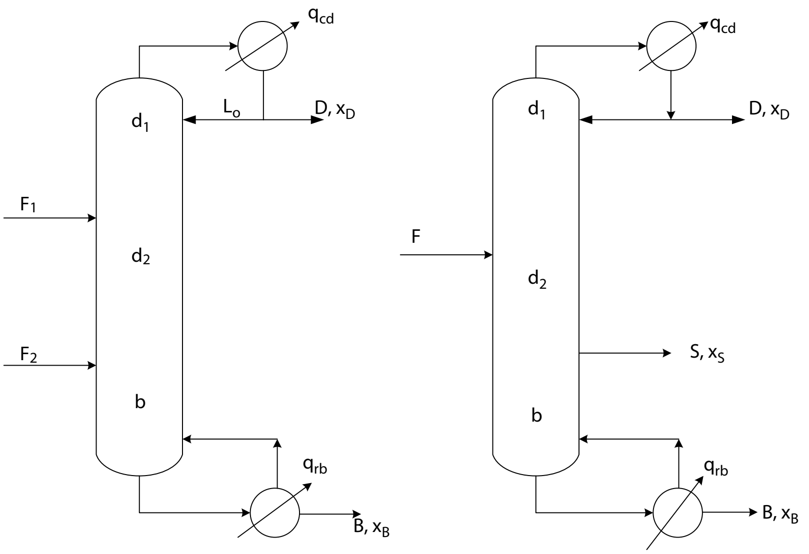


Figure 19.69 Ponchon analysis of a column with multifeed and with side stream.

of the line d_1 f′ and a vertical line at x_B. The difference point for the intermediate section d_2 is determined by the intersection of lines sb and fd_1. The number of equilibrium stages may be stepped off as usual.

The advantage of the H–x–y chart lies in the fact that the heat quantities required for the distillation are clearly indicated. Therefore, the higher the reflux ratio, the more heat must be removed per mole of product. This immediately shows that both q_c and Q_B are increased. The Ponchon–Savarit method is applicable to a wide range of binary flash and distillation problems. The method is more rigorous and may be used for a larger class of binary systems than the McCabe–Thiele method. However, the final choice between the Ponchon–Savarit and McCabe–Thiele methods must be based on the desired accuracy of results and the availability of data.

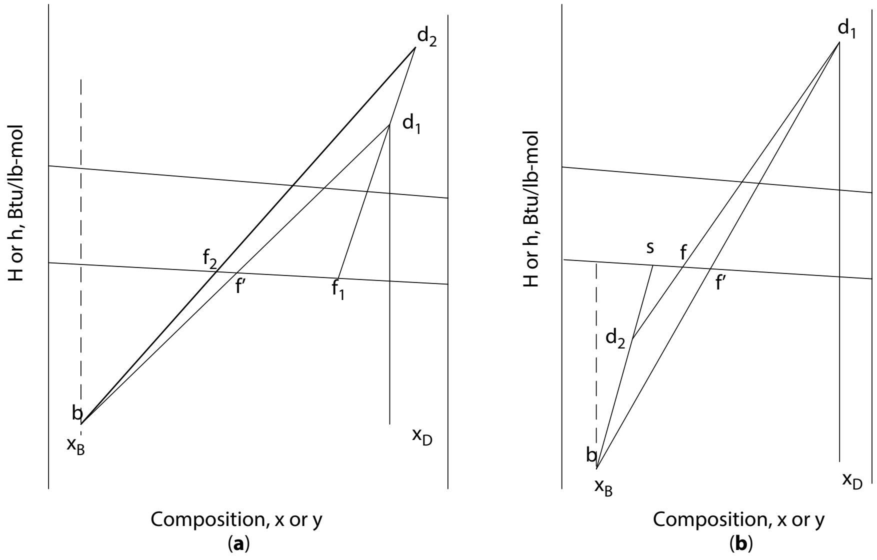


Figure 19.70 Ponchon analysis of a column with two feed streams.

Example 19.30: Ponchon Unequal Molal Overflow in Ammonia–Water System

An ammonia–water recirculating solution of 62 wt% is to be stripped of the ammonia for recovery by condensation at 260 psia with river water cooling. The overhead ammonia product is to be at least 99.5 wt%, and the bottoms should approach 0.05 wt% ammonia. The feed enters as a liquid at its boiling point, with an enthalpy of 42 Btu/lb.

Enthalpy Diagram

This is prepared by reading the h and H values from the Jennings and Shannon Aqua-Ammonia Tables [127] at 260 psia and various wt%'s of ammonia in the liquid. The tie lines connect the vapor compositions with the equilibrium liquid values (Figure 19.71).

Vapor–Liquid Equilibrium Diagram

Prepared from corresponding x and y values in Reference 35 at 260 psia (Figure 19.72).

Number of Trays

$x_F = 0.62$ weight fraction ammonia
$x_D = 0.995$ weight fraction ammonia
$x_B = 0.0005$ weight fraction ammonia

1. Minimum Reflux

$$\left(\frac{L}{D}\right)_{min} = \frac{(M_D)_{min} - H_D}{H_D - h_D} \tag{19.347}$$

From enthalpy–composition diagram:
$H_D = 590$ Btu/lb
$h_D = 92$ Btu/lb (assuming no subcooling)
$(M_D)_{min} = 596$ Btu/lb

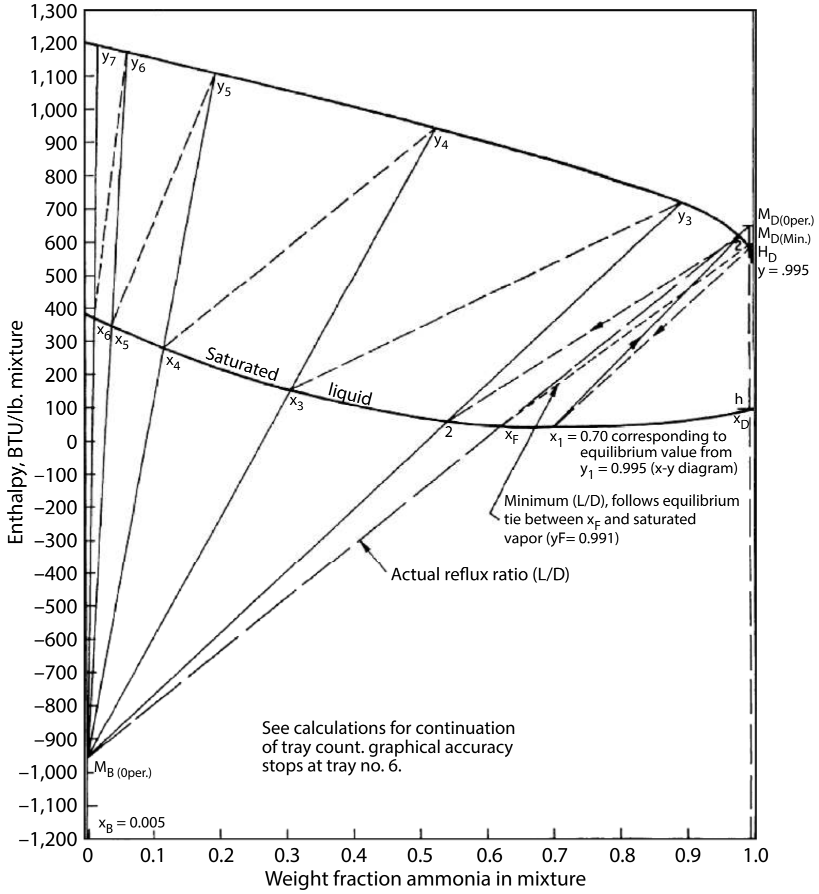


Figure 19.71 Ponchon type diagram for ammonia–water distillation of Example 19.27.

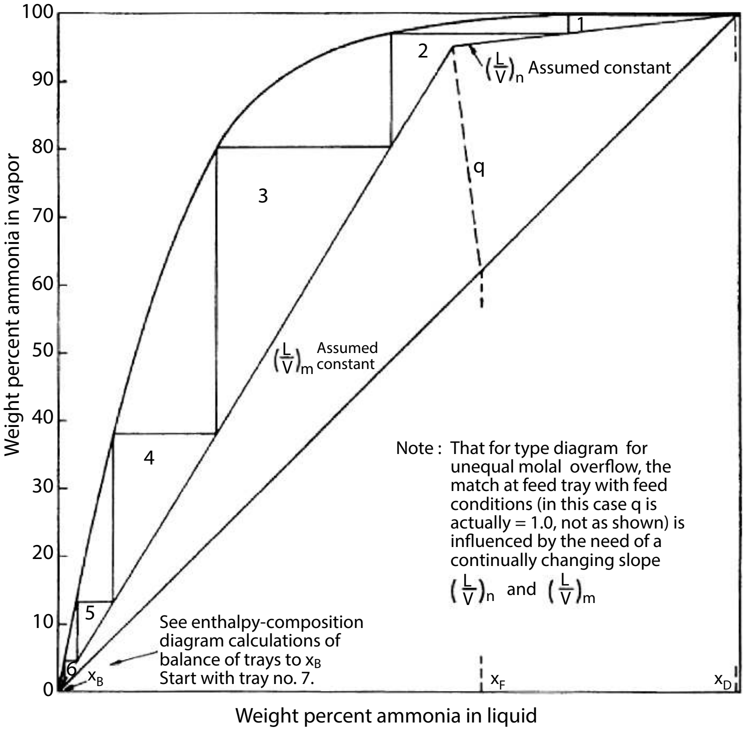


Figure 19.72 McCabe–Thiele diagram for ammonia–water system.

$(M_D)_{min}$ is determined by reading the equilibrium y value corresponding to the feed composition 0.62 from the x–y diagram, noting it on the enthalpy diagram on the saturated vapor curve, and connecting the tie line, then extending it on to intersect with the x_D ordinate 0.995, reading $(M_D)_{min}$ = 596 Btu/lb

$$\left(\frac{L}{D}\right)_{min} = \frac{596-590}{590-92} = 0.012$$

2. Operating reflux ratio, L/D

$$\text{Select } (L/D)_{actual} = 10\,(L/D)\text{min} = 10\,(0.012) = 0.12$$

This is not unusual to select an operating reflux ratio ten, or even fifty times such a low minimum. Selection of a higher reflux will result in a reduction in the number of trays required. As a consequence, this becomes a balance of the reduction in trays versus operating and capital expense in handling the increased liquid both external to the column and internally.

3. Operating M_D

$$\left(\frac{L}{D}\right)_{act} = 0.12 = \frac{M_D - 590}{590-92}$$

$$M_D = 649.8 \text{ Btu/lb}$$

Locate this value on the diagram and connect it to the feed point, x_F. Extend this line to intersect the bottoms condition ordinate (extended), x_B. In this case, it is impossible to represent the value, x = 0.005, accurately, but construct it as close as possible to the required condition. M_B is now located. The accuracy of this point will be improved later in the problem.

Following the procedures shown in Figures 19.65, 19.71, and 19.72, the trays are constructed from the top or overhead down toward the bottom. The x values are read to correspond to the y values constructed. This establishes the tie line. When the x value tie line points (representing the trays) cross the feed ordinate, the construction is shifted from using the point M_D to the point M_B. Note that only 1½ theoretical trays are required above the feed, since this is predominantly a stripping type operation. The number of theoretical trays or stages which can be easily plotted is six to seven counting down from the top. The sixth tray is too inaccurate to use graphically. Instead of calculating the balance of the trays assuming a straight-line equilibrium equation from tray seven to the end, the plot could be enlarged in this area and the trays stepped off. By reference to the x–y diagram, it can be seen that the equilibrium line from x = 0.02 to x = 0 is straight.

For the condition of straight operating and equilibrium curves, the number of plates can be calculated including the "reference" plate (number seven in this case) [85].

$$N_B = \frac{\ln\left[\frac{[(VK/L)-1]\left[(x_m/x_{1B})-1\right]}{(V/L)(K-1)}+1\right]}{\ln(VK/L)} \tag{19.348}$$

where

N_B = number of trays from tray m to bottom tray, but not including the still or reboiler
x_m = tray liquid mol fraction for start of calculations (most volatile component)
x_{1B} = mol fraction most volatile component in bottoms

For the lower end of the equilibrium curve,
$y_m = 5.0\, x_m$ (by slope calculation of x–y diagram)

For the stripping section: consider top seven trays, vapor entering tray No. 6, $y_7 = 0.02$, m = tray 7, m + 1 = tray 6, reading from diagram,

$$(L/V)_m = 1 - \frac{H_m - h_{m+1}}{M_B - h_{m+1}} = 1 - \frac{(1{,}190 - 369)}{(-960 - 369)} = 1.618$$

use $H_m = 1{,}190$
$h_{m+1} = 369$
$M_B = -960$

$$x_7 = \frac{y_7}{5.0} = \frac{0.02\,(\text{from graph})}{5.0} = 0.004$$

$$N_B = \frac{\ln\left[\dfrac{(5.0/1.618 - 1)0.004/0.0005 - 1}{(1/1.618)(5.0 - 1)}\right]}{\ln(5/1.618)}$$

N_B = 1.71 trays (theoretical) not including reboiler, but including tray number 7, the one used as reference.
Total trays = 7 (from diagram plus (1.71 − 1) = 7.7 theoretical, plus a reboiler or 8.7 including a reboiler.

Tray efficiency is calculated as previously described, but stripping tray efficiencies are normally lower than rectification efficiencies. For ammonia–water stripping, most overall efficiencies run 50–60%.

Note that if the problem of accurate graphical representation occurs in the rectification end of the diagram, the corresponding relationship to use in order to calculate the balance of the trays (assuming straight line operating and equilibrium lines in the region) is [85]

Rectifying section:

$$N_n = \frac{\ln\left[\dfrac{(1 - L/V) + x'_n/y'_D(L/V - K')}{1 - K'}\right]}{\ln(L/K'V)} - 1 \tag{19.349}$$

where
K' = equilibrium constant for the least volatile component, $K' = y/x$
N_n = number of plates above (but not including) reference plate n
y', x' = mole fractions least volatile component

Example 19.31: Ponchon–Savarit Unequal Molal Overflow in Heptane–Ethyl Benzene System

A saturated liquid feed mixture composed of 45 mole % heptane and 55 mole % ethyl benzene is to be fractionated at 760 mm Hg to produce a distillate containing 98% mole % heptane and a residue containing 99 mole % ethyl benzene. Using the enthalpy–concentration diagram based on molal enthalpy and mole fraction as shown in the table below, determine the following:

1. The minimum reflux ratio, L_o/D
2. The number of theoretical stages for an operating reflux ratio of $2.5R_{min}$.
3. The condenser duty, kcal/kg mole.
4. The reboiler duty, kcal/kg mole.

Solution

Temperature, °C	x_H	y_H	H_L, kcal/kg mole	H_V, kcal/kg mole
136.2	0	0	5.92	14.52
129.5	0.08	0.233	5.697	14.203
122.9	0.185	0.428	5.52	13.92
119.7	0.251	0.514	5.44	13.87
116.0	0.335	0.608	5.36	13.61
110.8	0.487	0.729	5.26	13.45
106.2	0.651	0.834	5.17	13.11
103.0	0.788	0.904	5.16	12.94
100.2	0.914	0.963	5.124	12.784
98.5	1.0	1.0	5.1	12.675

Figures 19.73a and 19.73b show the VLE and enthalpy composition diagrams of heptane, respectively, from the above data.

On the enthalpy–composition diagram, vertical lines are constructed at $x_B = 0.01$, $x_F = 0.45$ and $x_D = 0.98$. The feed point F is the intersection of the feed composition line and the liquid enthalpy curve. The point N is located by producing the line FD to the vertical through x_D. The point D on the vapor enthalpy curve corresponds to the value of y in equilibrium with $x_F = 0.45$ and is obtained from the equilibrium curve at y = 0.699.

1. The point N enables the minimum reflux ratio to be determined from

$$R_{\min} = \frac{\text{Length of NA}}{\text{Length of AB}}$$

And the length NB represents the heat removed in the condenser per unit mass of liquid at its boiling point. Points vertically above N correspond to actual reflux ratios (such as P_1), and thus the greater the value of R, the greater the

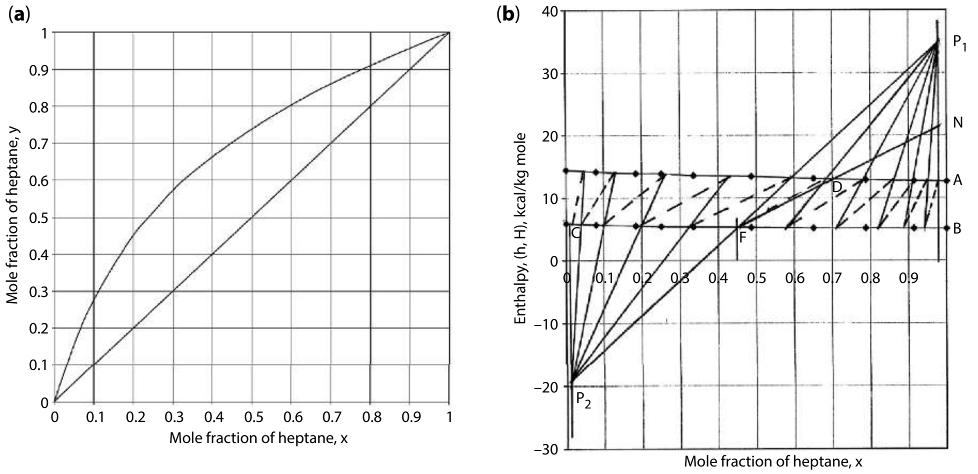


Figure 19.73 (a) VLE diagram of heptane–ethyl benzene mixture at 760 mm Hg. (b) Enthalpy composition diagram of heptane–ethyl benzene mixture at 760 mm Hg.

distance P_1N, and fewer number of plates are required for the separation. However, as P_1B represents the condenser duty, this also rises so that an economic selection of the reflux ratio must be made.

$$\begin{aligned} R_{min} &= \frac{\text{Length of NA}}{\text{Length of AB}} \\ &= (21.5-12.7)/(12.7-5.1) \\ &= 1.158 \end{aligned}$$

2. The number of theoretical stages for an operating reflux ratio of 2.5 R_{min} is determined as follows:

$$R = 2.5 \times 1.158 = 2.895$$

At the reflux ratio of 2.895, the point P_1 on Figure 19.73b is determined by

$$\frac{L}{D} = 2.895 = \frac{P_1A}{AB} = \frac{(P_1 - 12.7)}{(12.7 - 5.1)}$$

$$P_1 = 34.70 \text{ kcal/kg mole}$$

Hence P_1 has coordinates (0.98, 34.70)

The point P_2 is located by producing the line P_1F to its intersection with the vertical through x_B. The length P_2C represents the heat input to the reboiler per unit mass of bottom product. The coordinate of P_2 is -19 kcal/kgmole having the coordinate (0.01, -19). Points P_1 and P_2 are known as the difference points on the rectifying and stripping sections of the column. The construction of the tie lines may start from either side of Figure 19.73b, indicating either the conditions at the top or the bottom of the column and proceed as indicated for the rectifying and stripping sections of the column. In either case, when an equilibrium tie line crosses the line connecting the difference points (P_1P_2) through the feed condition, the other difference point is used to complete the construction. The number of tie lines indicates the theoretical number of stages.

From Figure 19.73b, the number of theoretical stages is 11.

3. The heat removed in the condenser per unit mass of product (condenser duty) in kcal/kg mole is determined from Figure 19.73b by length P_1B.

$$P_1B = 34.7 - 5.1 = 29.6 \text{ kcal/kg mole}$$

4. Heat required in the reboiler per unit mass of product is (reboiler duty) in kcal/kg mole is length P_2C in Figure 19.73b.

$$P_2C = 5.9 - (-19) = 24.9 \text{ kcal/kg mole}$$

Example 19.32: Ponchon–Savarit Unequal Molal Overflow in A-B System

A feed stream A and B consisting of boiling liquid and saturated vapor whose enthalpy is 1200 kJ/mol contains 40 mol % of A and is to be separated to produce a distillate of 99% of A and a bottom product containing not more than 2% of A. Calculate the following:

1. The minimum number of theoretical plates in the column to effect the separation.
2. The minimum reflux ratio, R_{min}.
3. The number of actual plates required if the column efficiency is 60% and the reflux ratio is $1.2R_{min}$.
4. The optimum feed plate for this condition.

Solution

Equilibrium and Enthalpy data.

Vapor–liquid equilibrium data

$_xA$	0	0.05	0.10	0.15	0.20	0.30	0.40	0.50
y_A	0	0.30	0.58	0.71	0.79	0.90	0.96	0.98

x_A	0.6	0.80	1.0
y_A	0.99	0.995	1

Enthalpy data (kJ/kmol)

x_A	0	0.20	0.40	0.60	0.80	0.90	0.99
H_L, kJ/kmol$_A$	767	418	198	116	163	209	279
H_V, kJ/kmol	2765	2580	2370	2160	1905	1765	1490

1. Figure 19.74 shows the enthalpy composition diagram of A, and the minimum number of theoretical plates occurs at total reflux when the point P_1 on $x = x_D$ is at infinity. The operating lines are therefore vertical. Further, the condition of the feed is located between the liquid and vapor enthalpy lines for a partly vaporized feed. For a feed at its boiling point, or as a saturated vapor, F lies on the liquid and vapor enthalpy lines, respectively. Similarly, for a liquid below its boiling point or as superheated vapor, the point F lies below or above the respective curves. With the composition and percent vapor specified, the point F may be located.
   For the minimum number of theoretical plates, starting at the top plate, the vapor leaving the plate has composition $y_T = 0.99$. The corresponding value of x_T from the equilibrium data is $x_T = 0.6$. From these points, the equilibrium tie line $x_T\ y_T$ is drawn. At total reflux, the operating line is vertical, and $y_{T-1} = 0.6$ and $x_{T-1} = 0.11$. y_{T-2} is located at $y_{T-2} = 0.11$ and x_{T-2} is found to equal $x_{T-2} = 0.02$. Therefore, two theoretical plates are required at total reflux.
2. The minimum reflux ratio is located by producing the tie line that passes through F to the line $x = x_D$ at N.

   Then

$$\begin{aligned} R_{min} &= \text{length NA/length AB} \\ &= (2880 - 1500)/(1500 - 290) \\ &= 1.14 \end{aligned}$$

3. $R = 1.2\ R_{min} = 1.368$. Hence, the difference points P_1 and P_2 are
   $P_1 = 3155$ kJ/mol, $P_2 = -80$ kJ/mol. The of theoretical stages are stepped–off as shown in Figure 19.74, and it is found that 5 ideal plates are required. For the column efficiency = 60%, the number of actual plates is 5/6.0 = 8 plates.
4. The feed tray is located on the diagram by the tie line that passes through the line P_1FP_2. The optimum feed plate is between the second and third stages.

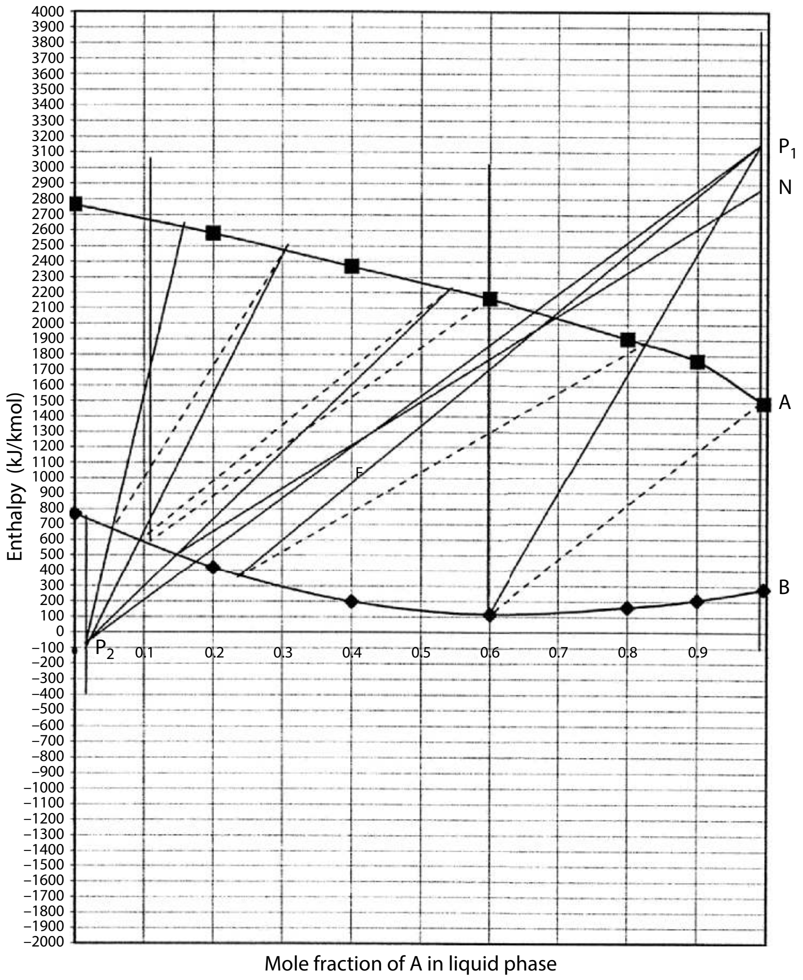


Figure 19.74 Enthalpy vs. composition of A-B system of Example 19.29.

19.37 Multicomponent Distillation

The basic background and understanding of binary distillation apply to a large measure in multicomponent problems. Reference should be made to Figure 19.1 for the symbols.

Multicomponent distillations are more complicated than binary systems due primarily to the actual or potential interaction of one or more components of the multicomponent system on the others. These interactions may be in the form of vapor–liquid equilibria, such as azeotrope formation, or chemical reaction, etc., which may affect the activity relations, and hence cause deviations from ideal behavior. For example, some systems are known to have two azeotrope combinations in the distillation column. Sometimes these can be "broken" by addition of a third chemical or hydrocarbon.

To properly handle the changing composition relationships, it is almost essential to use computational techniques if good accuracy is to be achieved. Even calculations with three component systems become tedious using calculators without significant internal memory. Computers can be programmed to easily handle the complexities of calculations involving techniques of trial and checking for convergence to a preset acceptable limit.

Digitally programmable algorithms for determining convergence are often the heart of an efficient multicomponent calculation. There are several techniques incorporated into many programs [128–133]. Several process simulators on the market can be readily employed to size multicomponent distillation/fractionation columns, and Table 19.28 shows the most commonly used commercial process simulators. Most of these programs can be licensed by colleges/universities for educational purpose at nominal cost.

Table 19.28 Simulation packages.

Name	Type	Source	Internet address http//www.
HYSY	Steady-state and dynamic	Aspen Technology Inc. Ten Canal Park Cambridge, MA 02141-2201, USA	Aspentech.com
Aspen Plus	Steady-state	Aspen Technology Inc. Ten Canal Park Cambridge, MA 012141-2201, USA	Aspentech.com
CHEMCAD	Steady-state	Chemstations, Inc. 2901 Wilcrest Suite 305 Houston, TX 77042	Chemstation.net
PRO/11 and DYNSIM	Steady-state and dynamic	SimSci-Esscor 5760 Fleet Street Suite 100 Carlsbad, CA	Simsci.com
DESIGN II	Steady-state	WinSim Inc. P.O. Box 1885	Winsim.com
UniSim Design	Steady-state and dynamic	Honeywell 300-250 York Stree London, Ontario N6A 6K2, Canada	Honeywell.com
ProSim	Steady-state and dynamic	ProSim SA Stratege Batiment A BP 27210 F-31672 Labege Cedex, France	prosim.net

Note: Contact the companion website to check the full features of the most recent versions of the programs.

Key Components

The two components in a feed mixture whose separations will be specified.

1. Adjacent keys: key components that are adjacent with respect to their volatilities.
2. Split keys: key components that are separated in volatilities by a non-key component, i.e., the system of components contains one or more whose volatilities fall between the volatilities of the designated keys.
3. Light key: the designation of the key component with the highest volatility of the two key components.
4. Heavy key: the designation of the key component with the lowest volatility of the two key components.
5. Example: component designations

Component	Relative Volatility $\alpha_{l/h}$–7°F. and 550 psia	Designation
Hydrogen	11.7	Lighter than Key
Methane	3.7, α_l	*Light Key, 1*
Ethylene	1.0, α_h	*Heavy Key, h*
Ethane	0.72	Heavier than Key
Propylene	0.23	Heavier than Key
Propane	0.19	Heavier than Key

Hengstebeck [134] presents a simplified procedure for reducing a multicomponent system to an equivalent binary using the "key" components. From this, the number of stages or theoretical plates and reflux can be determined using conventional binary procedures and involving the McCabe–Thiele method.

Liddle [94] presents a shortcut technique for multicomponent calculations based on improving the Fenske and Gilliland correlations.

19.37.1 Minimum Reflux Ratio-Infinite Plates

This is the smallest value of external reflux ratio (L/D), which can be used to obtain a specified separation. This is not an operable condition. Knowledge of this minimum reflux ratio helps to establish an economical and practical operating ratio. Ratios of 1.2 to 2.0 times the minimum are often in the economical range for hydrocarbon systems. However, as already discussed, high reflux rates increase column size (but reduce number of trays required), reboiler size, steam rate, condenser size, and coolant rate.

For adjacent key systems, all components lighter than the light key appear only in the overhead, and all components heavier than the heavy key appear only in the bottoms, and the keys each appears in the overhead and bottoms at levels in accordance with specifications.

For a split key system, the lights and heavies distribute the same way as for adjacent key systems. However, the component(s) between the keys also distribute to overhead and bottoms.

At minimum reflux, the regions in which the number of trays approaches infinity (called the pinch zones and region of constant compositions) are:

1. Binary system: pinch zone adjacent to feed plate
2. Multicomponent:
    a. Three components with no component lighter than light key: pinch zone in stripping section adjacent to feed plate.
    b. Three components with no component heavier than heavy key: pinch zone in rectifying section adjacent to feed plate.
    c. Three components mixture: pinch zones may be above and below feed plate.
    d. Greater than four components: pinch zones appear in rectifying and stripping sections.

For systems with one side stream draw-off, either above or below the feed, Tsuo *et al.* [135] propose a method for recognizing that the minimum reflux ratio is greater for a column with side stream draw-off. At the side stream, the operating line has an inflection. For multifeed distillation systems, the minimum reflux is determined by factoring together the separate effect of each feed [136].

Lessi [137] proposes a detailed graphical procedure for figuring multicomponent minimum reflux by a graphical extension of a McCable–Thiele diagram, assuming infinite plates or equilibrium stages. In this traditional model, the concentration in the distillate of the components heavier than the heavy key is assumed to be zero, and the heavy key component reaches its maximum concentration at the upper pinch point (see Figures 19.48 and 19.50). Therefore, the assumption is that only the heavy and light keys are present at the upper pinch point, similar in concept to the handling of a binary mixture [138]. The method assumes (a) only the key components are distributed, (b) no split key components exist, and (c) total molal overflow rates and relative volatilities are constant. This method provides good agreement with the detailed method of Underwood.

Yaws *et al.* [139] provide an estimation technique for recovery of each component in the distillate and bottoms from multicomponent distillation using short-cut equations and involving the specification of the recovery of each component in the distillate, the recovery of the heavy key component in the bottoms, and the relative volatility of the light key component. The results compare very well with plate-to-plate calculations (Figure 19.75) for a wide range of recoveries of 0.05 to 99.93% in the distillate.

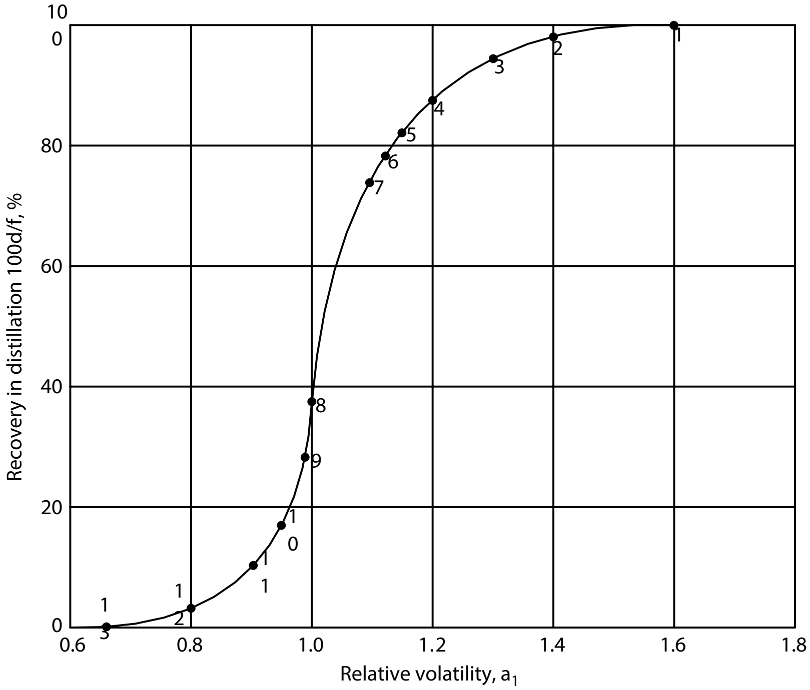


Figure 19.75 Yaws short-cut method compared to plate-to-plate calculations. (Used by permission, Yaws, C. L., *et al.* Hydrocarbon Processing, V. 58, No. 2, pp 99, (1979). Gulf Publishing Co., All rights reserved).

Equations Surveyed

The availability of desktop computers and simulation software packages has permitted optimization of the number of equilibrium stages in multipurpose fractionating towers and absorbers. Vigorous iterative plate-to-plate computation is still used to study a wide range of process conditions. However, preliminary design with graphical correlations often helps to arrive at an approximate optimum number of stages before reverting to established design methods. Figure 19.1 shows a schematic of a fractionating column with two or more multicomponent and associated equipment items. Alternatives to the preliminary design approach are the short-cut methods that estimate a realistic optimum number of theoretical stages. Generally, short-cut methods were developed to establish the minimum number of stages at total reflux, and also the minimum reflux at an infinite number of theoretical stages. Many authors have correlated the minimum reflux and minimum stages with operating reflux and corresponding theoretical equilibrium stages required. These methods have enabled the reflux ratio and number of plates to be chosen to achieve an optimum relationship based on investment and operating costs.

Mathematical Modeling

The distribution of components between the distillate and bottoms is given by the Hengstebeck-Geddes equation [139–141]:

$$\log\left[\frac{d_i}{b_i}\right] = A + B\log\alpha_i \tag{19.350}$$

where

d_i = moles of component i in the distillate
b_i = moles of component i in the bottoms.
α_i = relative volatility of component i
A, B = correlation constants.

A material balance for the ith component in the feed is

$$f_i = d_i + b_i \tag{19.351}$$

The quantity of component i in the distillate can be expressed as mole fraction recovered or d_i/f_i. Alternatively, in the bottoms, the mole fraction of component i recovered is b_i/f_i.

If Equation (19.350) is expressed with respect to the heavy key component, then

$$\log\left[\frac{d_{HK}}{b_{HK}}\right] = A + B\log\alpha_{HK} \tag{19.352}$$

The relative volatility of the heavy key component (i.e., $\alpha_{HK} = 1.0$).

Therefore,

$$A = \log\left[\frac{d_{HK}}{b_{HK}}\right] \tag{19.353}$$

but

$$f_{HK} = d_{HK} + b_{HK} \tag{19.354}$$

and

$$f_{LK} = d_{LK} + b_{LK} \tag{19.355}$$

Therefore,

$$A = \log\left[\frac{f_{HK} - b_{HK}}{b_{HK}}\right] \tag{19.356}$$

Equation (19.356) can be expressed in terms of mole fraction recovered as

$$A = \log\left[\frac{(1.0 - b_{HK}/f_{HK})}{(b_{HK}/f_{HK})}\right] \tag{19.357}$$

Substituting Equation (19.353) into Equation (19.350) and expressing in terms of the light key component,

$$\log\left[\frac{d_{LK}}{b_{LK}}\right] = \log\left[\frac{d_{HK}}{b_{HK}}\right] + B\log\alpha_{LK} \tag{19.358}$$

Therefore,

$$\begin{aligned} B &= \frac{\log\left[\left(\dfrac{d_{LK}}{b_{LK}}\right)\Big/\left(\dfrac{d_{HK}}{b_{HK}}\right)\right]}{\alpha_{LK}} \\ &= \frac{\log\left[\left(\dfrac{(d_{LK}/f_{LK})}{\left(\dfrac{b_{LK}}{f_{LK}}\right)}\right)\left(\dfrac{d_{HK}}{b_{HK}}\right)\right]}{\log\alpha_{LK}} \end{aligned} \tag{19.359}$$

Expressing Equation (19.359) in terms of fractional recoveries

$$B = \frac{\log\left[\left(\frac{d_{LK}/f_{LK}}{\frac{f_{LK}-d_{LK}}{f_{LK}}}\right)\left(\frac{b_{HK}/f_{HK}}{\frac{f_{HK}-b_{HK}}{f_{HK}}}\right)\right]}{\log\alpha_{LK}} \tag{19.360}$$

or

$$B = \frac{\log\left[\left(\frac{d_{LK}/f_{LK}}{1-\frac{d_{LK}}{f_{LK}}}\right)\left(\frac{b_{HK}/f_{HK}}{1-\frac{b_{HK}}{f_{HK}}}\right)\right]}{\log\alpha_{LK}} \tag{19.361}$$

The recoveries of the ith component in the distillate and bottoms:

Equations (19.350) and (19.351) are

$$\log\left[\frac{d_i}{b_i}\right] = A + B\log\alpha_i \tag{19.362}$$

$$f_i = d_i + b_i \tag{19.351}$$

Expressing Equations (19.362) and (19.351) in terms of the recoveries of the ith component.

That is,

$$\log\left[\frac{d_i / f_i}{1-d_i / f_i}\right] = A + B\log\alpha_i \tag{19.363}$$

Hence,

$$\begin{aligned}\left[\frac{d_i/f_i}{1-d_i/f_i}\right] &= 10^A \cdot 10^{\left(\log_{10}\alpha_i^B\right)} \\ &= 10^A \cdot \alpha_i^B\end{aligned} \tag{19.364}$$

$$\frac{d_i}{f_i} = 10^A \cdot \alpha_i^B\left(1-\frac{d_i}{f_i}\right) \tag{19.365}$$

The recovery of the ith component in the distillate is given by

$$\frac{d_i}{f_i} = \frac{\left(10^A \cdot \alpha_i^B\right)}{\left(1+10^A \cdot \alpha_i^B\right)} \tag{19.366}$$

and the recovery of the ith component in the bottoms is

$$\frac{b_i}{f_i} = 1-\frac{d_i}{f_i} \tag{19.367}$$

$$=1-\frac{\left(10^{A}\cdot\alpha_i^{B}\right)}{\left(1+10^{A}\cdot\alpha_i^{B}\right)} \tag{19.368}$$

$$\frac{b_i}{f_i}=\frac{1}{\left(1+10^{A}\cdot\alpha_i^{B}\right)} \tag{19.369}$$

The correlation constants required for Equations (19.366) and (19.369) are obtained by specifying a desired recovery of the light key component LK in the distillate and the recovery of the heavy key component HK in the bottoms. The constants are calculated as follows:

$$A=-\log\left[\frac{(b_{HK}/f_{HK})}{1-(b_{HK}/f_{HK})}\right] \tag{19.370}$$

or

$$A=\log\left[\frac{1-(b_{HK}/f_{HK})}{(b_{HK}/f_{HK})}\right] \tag{19.371}$$

$$B=\frac{\log\left[\left(\frac{(d_{LK}/f_{LK})}{1-(d_{LK}/f_{LK})}\right)\left(\frac{(b_{HK}/f_{HK})}{1-(b_{HK}/f_{HK})}\right)\right]}{\log\alpha_{LK}} \tag{19.372}$$

19.37.2 The Fenske's Method for Total Reflux [142]

Fenske's [142] equation for determining the minimum equilibrium stages at total reflux was based on an ideal mixture. This suggests that the ratio of vapor pressures or the ratio of equilibrium vaporization of the key components is constant over the range of temperatures (that is, the relative volatilities are constant). Fenske expressed the minimum number of equilibrium stages as

$$N_{min}=\frac{\log\left[\left(\frac{x_{LK}}{x_{HK}}\right)_D\left(\frac{x_{HK}}{x_{LK}}\right)_B\right]}{\log\left[\frac{\alpha_{LK}}{\alpha_{HK}}\right]} \tag{19.183}$$

19.37.3 The Gilliland Method for Number of Equilibrium Stages [90]

The number of theoretical equilibrium stages required for a given separation at a given reflux ratio is often determined by empirical correlations [90, 143]. The abscissa X represents a reflux function as

$$X=\frac{R-R_m}{R+1} \tag{19.223}$$

Correspondingly, the ordinate Y represents a stage function as given by

$$Y = \frac{(N - N_m)}{(N+1)} \tag{19.221}$$

The ratios used for the axes of abscissa and ordinate were chosen because they provide fixed end points for the curve (X = 0.0, Y = 1.0 and at X = 1.0, Y = 0.0). The two functions were found to give good correlations. These correlations are shown in Figures 19.49a and 19.76. Gilliland's correlation has produced relevant results that offer the following advantages that:

- They represent an optimum solution concerning the location of the feed plate.
- The splitting for the two key components is verified.
- The maximum deviation using Gilliland's correlation in terms of tray number is within the 7% range [90].

The Gilliland correlation tends to be conservative for feeds with low values of the thermal condition of the feed (q). It can give inaccurate results when there is a large difference in tray requirements above and below the feed. The correlation is adequate for use in preliminary designs before a detailed analysis is carried out, but great caution should be exercised if it is used for final designs. Hines and Maddox [59] have described an improved correlation for determining the number of theoretical stages.

19.37.4 Underwood's Method [88, 144]

If an infinite or nearly infinite number of equilibrium stages are involved, a zone of constant composition must exist in the fractionating column. In this instance, there is no measurable change in the composition of liquid or vapor from stage to stage. Under these conditions, the reflux ratio can be defined as the minimum reflux ratio, R_{min}, with respect to a given separation of two *key* components (that is, the light key and heavy key) [144].

Therefore, for component i in the distillate

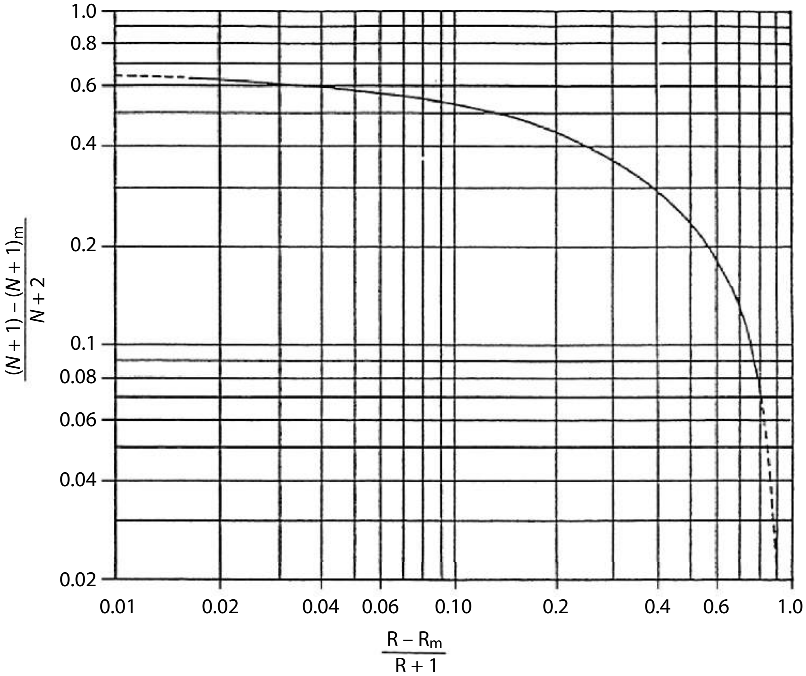


Figure 19.76 Gilliland correlation of reflux ratio, theoretical stages, minimum reflux ratio, and minimum number of stages [Source: Gilliland, Ind. Eng. Chem., 32:1101 (1940)]. $(N+1)_m$ = The number of theoretical stages at total reflux; Rm = The minimum reflux ratio; R = Reflux ratio; N = Number of theoretical stages.

$$R_{min} + 1 = \sum_{i=1}^{n} \frac{(\alpha_i - x_{i,D})}{(\alpha_i - \theta)} \tag{19.373}$$

where θ is Underwood's constant (or root of the equation), and must lie between the relative volatilities of the light and heavy keys (α_{LK} and α_{HK}). The number of components is n.

Correspondingly, for the feed

$$1 - q = \sum_{i=1}^{n} \frac{(\alpha_i \cdot x_{i,F})}{(\alpha_i - \theta)} \tag{19.374}$$

Equation (19.374) shows the relationship for the feed, where q is the fraction of feed that is liquid at the feed tray temperature and pressure. For a bubble point feed, q = 1, and for a dew point feed, q = 0. The minimum reflux ratio is determined from Equation (19.374) by substituting into Equation (19.373). This author [145] has developed a numerical method for computing θ and R_{min}, respectively. However, other methods should be tried if R_{min} gives a negative value. Also, it may be that the separation between the feed and the overhead can be accomplished in less than one equilibrium stage.

19.36.5 Equations for Describing Gilliland's Graph

Many equations have been proposed to describe Gilliland's curve for multicomponent distillation. However, the difficulty with some of these equations has been in meeting the end conditions of X = 0, Y = 1 and X = 1, Y= 0. A review of the many equations proposed by these authors is as follows:

Proposed Equations

1. Hengstebeck [138] (1961).

$$\log Y = A + B(\log X) + C(\log X)^2 + D(\log X)^3 + E(\log X)^4 \tag{19.375}$$

where
A = -1.3640187
B = -3.0920489
C = -3.407344729
D = -1.74673876
E = -0.33268897

2. Liddle [94] (1968)

$$\text{For } 0.0 \le X \le 0.01,\ Y = 1.0 - 18.5715X \tag{19.376}$$

$$\text{For } 0.01 \le X \le 0.9,\ Y = 0.545827 - 0.591422X + 0.002743/X \tag{19.377}$$

$$\text{For } 0.09 \le X \le 1.0,\ Y = 0.16595 - 0.16595X \tag{19.378}$$

3. van-Winkle and Todd [132] (1971)

$$\text{For } 0.0078 < X < 0.125,\ Y = 0.5039 - 0.5968X - 0.0908 (\log X) \tag{19.379}$$

$$\text{For } 0.125 < X = 1.0,\ Y = 0.6257 - 0.9868X + 0.516X^2 - 0.1738X^3 \tag{19.380}$$

4. Molokavov [146] (1972)

$$0 \le (X, Y) \le 1.0$$
$$Y = 1 - \exp\left[\left(\frac{1 + 54.4X}{11 + 117.2X}\right)\left(\frac{X - 1}{X^{0.5}}\right)\right] \tag{19.381}$$

5. Hohman and Lockhart [147] (1972)

$$\begin{gathered} X = 0,\ Y = 0.65 \\ X = 1.0,\ Y = 0.067 \\ Y = \frac{0.65 - 0.50X}{1.0 + 1.25X} \end{gathered} \tag{19.382}$$

6. Eduljee [91] (1975)

$$X = 0,\ Y = 1.0$$

$$X = 1.0,\ Y = 0$$

$$Y = 0.75 - 0.75X^{0.5668} \tag{19.383}$$

7. Huan-Yang Chang [148] (1985)

$$\text{For } X = 0,\ Y = 1.0$$

$$X = 1.0, Y = 0$$
$$Y = 1 - \exp\left(1.49 + 0.315X - \frac{1.805}{X^{0.1}}\right) \tag{19.384}$$

8. Harg [149] (1985)

$$\text{For } X = 0,\ Y = 1.0$$

$$X = 1.0,\ Y = 0$$

$$Y = 1 - X^{1/3} \tag{19.385}$$

9. McCormick [96] (1988)

$$X = 0,\ Y = 1.0$$

$$X = 1.0,\ Y = 0$$

$$Y = 1 - X^{B}$$

where

$$B = 0.105(\log X) + 0.44 \tag{19.386}$$

10. Rusche [150] (1999)

$$Y = 0.2788 - 1.3154X + 0.4114X^{0.291} + 0.8268 \ln X + 0.9020 \ln\left(X + \frac{1}{X}\right) \quad (19.387)$$

where

$$X = \frac{R - R_{min}}{R + 1}, \; Y = \frac{N - N_m}{N + 1} \quad (19.388)$$

From which the number of theoretical trays is

$$N = \frac{N_m + Y}{1 - Y} \quad (19.389)$$

The Gilliland correlation is conservative for feeds with low values of q (thermal condition of the feed) and can result in erroneous results when there is large difference in tray requirements above and below the feed. The various correlations illustrated are for preliminary exploration of design variables, which should be further refined using simulation packages (Aspen, Chemcad, Hysys, UniSim®, Pro II, etc.). These correlations should be employed with caution, if applied to final design.

From the equations listed, those of McCormick give good agreement in the normal operating range of real towers. The reflux ratio R is calculated as a multiple of the minimum reflux ratio, R_{min}.

That is,

$$R = \text{FACTOR} \cdot R_{min} \quad (19.390)$$

19.37.6 Operating Reflux Ratio, R

As the reflux ratio increases above the minimum, the number of trays required for a given separation decreases; hence, the capital cost of the column decreases. However, increasing the reflux ratio will increase the vapor rate in the tower, and this will impact the cost of the reboilers and condensers, as well as increasing cooling water and steam costs. Hence, there is an optimum reflux ratio for any specified separation as illustrated later in this volume. Experience has shown that the value of this optimum normally falls in the range $1.03 < R/R_{min} < 1.3$. For the first estimate, the multiplier "FACTOR" generally varies from 1.2 to about 1.5 for conventional columns, but because of economics, the range is now between 1.05 and 1.20.

Rule of Thumb

The operating reflux ratio is chosen so that $R/R_{min} = 1.2$.

19.37.7 Feed Tray Location

The approximate location can be determined by the ratio of the total number of theoretical stages above and below the feed plate from the Fenske total reflux relation:

$$\frac{S_r}{S_s} = \frac{n+1}{m+1} = \frac{\log(x_l/x_h)_D \; (x_h/x_l)_F}{\log(x_l/x_h)_F \; (x_h/x_l)_B} \quad (19.391)$$

The relation is solved for S_r/S_s. The results are not exact, because the feed tray composition is very seldom the same as the feed, which is the assumption in this relation. Actually, the feed point or correct location for the feed may be wrong by two or three theoretical trays. This will vary with the system. It does mean, however, that when this approach is used for feed plate location, alternative feed nozzles should be installed on the column to allow for location of the best feed point to be determined by experiment. These extra nozzles are usually placed on alternate trays (or more) both above and below the calculated location. A minimum of three alternate nozzles should be available.

When the feed point is located by tray-by-tray calculation, the correct point can be established with greater confidence, but still alternate nozzles are suggested since even these detailed calculations can be inaccurate.

The actual number of trays in the rectifying section $(N_{act})_r$ can be determined by

$$S_M = S_r + S_s \tag{19.392}$$

$$S_M = S_s(S_r/S_s) + S_s \tag{19.393}$$

Solve for S_s because S_M and S_r/S_s are known.
Obtain S_r by difference.
$(N_{act})_r = S_r/E_o$ for total condenser; if partial condenser use

$$(S_r - 1)/E_o \tag{19.394}$$

$$(N_{act})_s = (S_s - 1)/E_o \text{ (for columns with reboilers)} \tag{19.395}$$

For systems with wide variation in relative volatility, the suggestion of Cicalese *et al.* [151] is often used to evaluate the theoretical total equilibrium stages in the rectifying and stripping sections:

$$S_r = \frac{\log \dfrac{(x_l/x_h)_D}{(x_l/h_h)_F}}{\log \alpha(\text{average above feed})} \tag{19.396}$$

$$S_s = \frac{\log \dfrac{(x_l/x_h)_F}{(x_l/h_h)_B}}{\log \alpha(\text{average above feed})} \tag{19.397}$$

where
S_r = number of theoretical trays/plates in rectifying section
S_s = number of theoretical trays/plates in stripping section

Maas [152] presents a useful analysis for selecting the feed tray in a multicomponent column. For accuracy, it involves the use of a tray-by-tray computer calculation.

19.37.8 Kirkbride's Feed Plate Location [153]

After the minimum number of stages and the minimum reflux ratio have been determined, the number of theoretical stages is then calculated. The ratio of the number of plates above the feed stage (including the partial condenser) to the number below the feed stage (including the reboiler) can be obtained using Kirkbride's empirical equation [153].

The equation was developed on the basis that the ratio of rectifying trays to stripping trays depends on the following:

- The fraction of the heavy key component (in the feed) removed in the overhead.
- The fraction of the light key component removed in the bottoms.
- The concentration of the heavy key component in the overhead.
- The concentration of the light key component in the bottoms.

Kirkbride's feed plate equation is expressed as

$$\log\left[\frac{m}{p}\right] = 0.206\log\left\{\left(\frac{B}{D}\right)\left(\frac{x_{HK}}{x_{LK}}\right)_F\left[\frac{(x_{LK})_B}{(x_{HK})_D}\right]^2\right\} \tag{19.398}$$

$$\frac{m}{p} = \left\{\left(\frac{B}{D}\right)\left(\frac{x_{HK}}{x_{LK}}\right)_F\left[\frac{(x_{LK})_B}{(x_{HK})_D}\right]^2\right\}^{0.206} \tag{19.399}$$

Akashah *et al.* [154] presented a modified version of Equation (19.398) or (19.399), which gives

$$m = p[\text{calculated From Eq. 19.399}] - 0.5\log N \tag{19.400}$$

$$N = m + p \tag{19.401}$$

where

B = molar flow of bottoms product
D = molar flow of top product
N = total number of theoretical stages
m = number of theoretical stages above the feed plate, including any partial condenser
p = number of theoretical stages below the feed plate, including the reboiler.
$(x_{LK})_F$ = concentration of the light key in the feed
$(x_{HK})_F$ = concentration of the heavy key in the feed
$(x_{LK})_B$ = concentration of the light key in the bottoms product
$(x_{HK})_D$ = concentration of the heavy key in the distillate product

Example 19.33: Multicomponent Distillation by Yaw's Method [139]

Assume a multicomponent distillation operation has a feed whose component concentration and component relative volatilities (at the average column conditions) are as shown in Table 19.29. The desired recovery of the light key component O in the distillate is to be 94.84%. The recovery of the heavy key component P in the bottoms is to be 95.39%.

The recoveries of the non-key components are estimated by first calculating the correlation constants:

$$b_{HK}/f_{HK} = 0.9539\text{, given}$$

$$A = -\log\left(\frac{0.9539}{1-0.9539}\right) = -\log 20.69 = 1.3158$$

$$d_{LK}/f_{LK} = 0.9484\text{, given}$$

Table 19.29 Yaw's method for selected distillation recovery from a specific feed for Example 19.33.

Component	f_i	α_i
M	0.10	2.30
N	0.13	1.75
O (LK)	0.25	1.45
P (HK)	0.23	1.00
Q	0.15	0.90
R	0.08	0.83
S	0.06	0.65

Used by permission, *Hydrocarbon Processing*, Yaw's, C. L., *et al.* V. 58 No. 2 (1979), p. 99, Gulf Pub. Co., all rights reserved.

$$\alpha_{LK} = 1.45, \text{ from Table 19.29}$$

$$B = \frac{\log\left[\left(\frac{0.9484}{1-0.9484}\right)\left(\frac{0.9539}{1-0.9539}\right)\right]}{\log 1.45}$$

$$= (\log 380.3)/(\log 1.45) = 15.988$$

The recovery of component M in the distillate is then

$$d_M/f_M = (10^{-1.3158}2.30^{15.988})/(1 + 10^{-1.3158}2.30^{15.988}) = 0.99997, \text{ from Equation (19.366).}$$

The recovery of component M in the bottoms is

$$d_M/f_M = 1/(1 + 10^{-1.3158}2.30^{15.988}) = 0.00003, \text{ from Equation (19.369).}$$

Repeating Equations (19.366) and (19.369) for each of the other non-key components in the feed mixture gives the results shown in Table 19.30. Good agreement was demonstrated. Table 19.30 also provides snapshot of the Excel spreadsheet calculations (Example 19.33.xlsx).

Computer Program

The author [145] has developed the Excel spreadsheet program (Example 19.33.xlsx) for short-cut multicomponent distillation calculations based on a series of equations, developed by Hengstebeck-Geddes [134, 139–141], while the minimum tray calculations (at total reflux) use the equations developed by Fenske [142]. The number of theoretical equilibrium stages is determined using Gilliland's [90] empirical correlations. The program has an option of entering either the equilibrium constant or the relative volatility with the feed stream key components.

It also arranges the relative volatilities in order of decreasing magnitude before proceeding to compute the correlation constants. The following assumptions are used in the program.

- There is only one feed stream with 12 or fewer key components.
- There is only one heavy key component.
- Feed components may be arranged in order of decreasing relative volatility, but the light key (LK) and heavy key (HK) components must be adjacent to each other.

Table 19.30 Results for Example 19.33 for multicomponent distillation.

Component	Percent recovery	
	In distillate (100 d_i/f_i)	**In bottoms (100 d_i/f_i)**
M	99.997	0.003
N	97.731	2.269
O (LK)	94.840	5.160
P (HK)	4.610	95.390
Q	0.889	99.111
R	0.245	99.755
S	0.005	99.995

Used by permission, *Hydrocarbon Processing*, Yaw's, C. L., *et al* V. 58 No. 2 (1979), p. 99, Gulf Pub. Co., all rights reserved.

- The feed enters the column at the optimum stage.
- The column produces two products (distillate and bottoms) with overhead condenser and bottom reboiler.

The program uses the bisection method to determine the Underwood's constant θ, or the root of Underwood's equation, and then proceeds to evaluate the minimum reflux ratio. From a given multiplier (FACTOR), the actual reflux ratio is calculated.

From the various equations that have been proposed to describe Gilliland's curve for multicomponent distillation, the author uses that of McCormick [Equation (19.386)] because it gives a good agreement in the normal operating range of real towers. The program then proceeds to determine the actual reflux ratio from the minimum reflux ratio using Equation (19.390). Using these equations with Equation (19.399), the program calculates the number of theoretical plates and the optimum location of the feed plate in a distillation column.

NB: This program is best used in preliminary design before a detailed design by a process simulator package is explored.

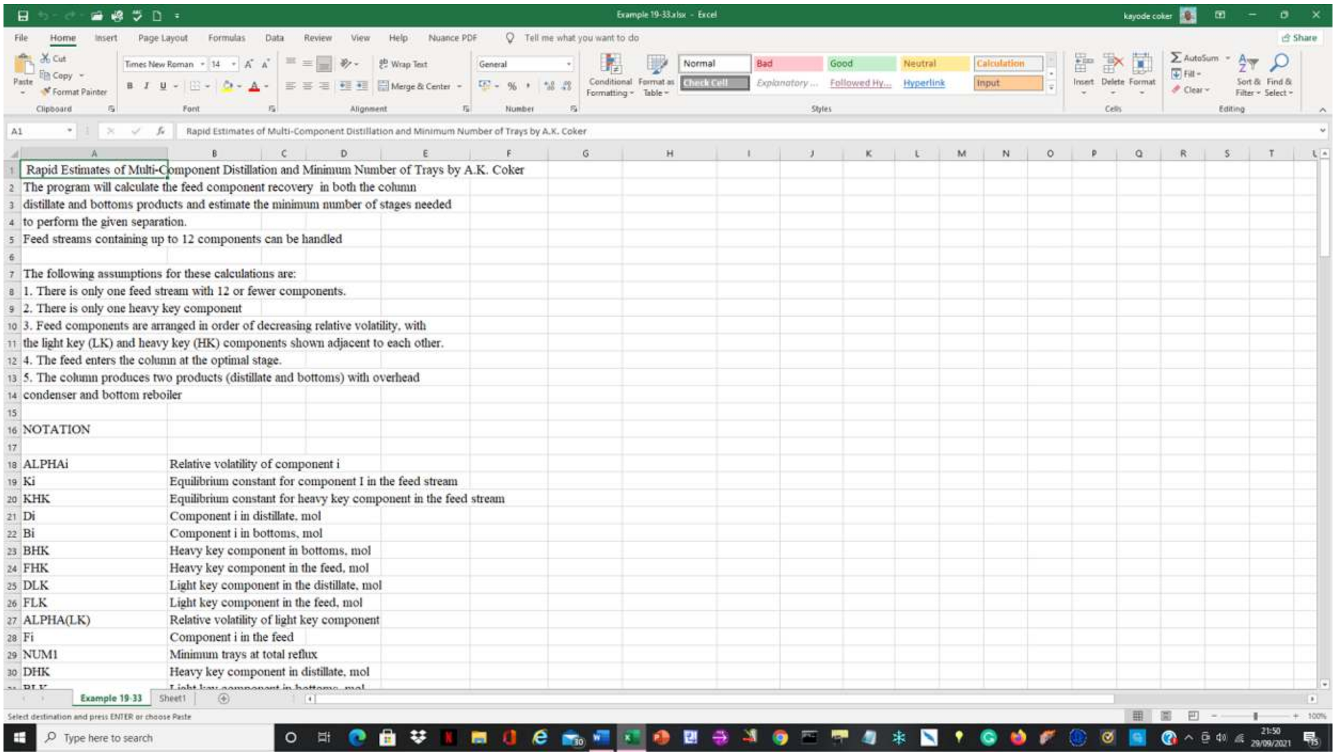
Table 19.31 shows snapshots of the Excel spreadsheet calculations (Example 19.33.xlsx) with the input data and results. The spreadsheet calculations determine, θ, the root of Underwood's equation, and the minimum reflux ratio R_{min} for multicomponent distillation. The program uses Microsoft Excel optimization technique such as Go-Seek or Solver to determine the Underwood's constant θ for a given feed condition (q), and then proceeds to calculate the minimum reflux ratio for various values of the distillate composition. The actual reflux ratio is determined by multiplying the minimum reflux ratio by a factor (FACTOR) of 1.15 to 1.3. The program uses the Fenske equation to determine the minimum number of plates at total reflux, using the developed correlation of Erbar–Maddox (Figure 19.77 or 19.77a). The program further calculates the number of theoretical plates and compares the value with the Rusche correlation (Equation 19.387). Figures 19.78a and 19.78b show snapshots of the calculations. The results in Table 19.30 by Yaws *et al.* [139] and in Table 19.31 from the Excel spreadsheet program (Example 19.33.xlsx) are in good agreement.

19.37.9 Algebraic Plate-to-Plate Method

Like any plate-to-plate calculation, this is tedious, and in most instances does not justify the time spent, because shorter methods give reasonably acceptable results. van Winkle [15] outlines the steps necessary for such calculations.

With current computer technology, there are several commercial programs available (as well as personal and private ones) that perform tray-to-tray stepwise calculations up or down a column, using the latest vapor pressure, K-values,

Table 19.31 Excel spreadsheet calculations (Example 19.33.xlsx) of Example 19.33.

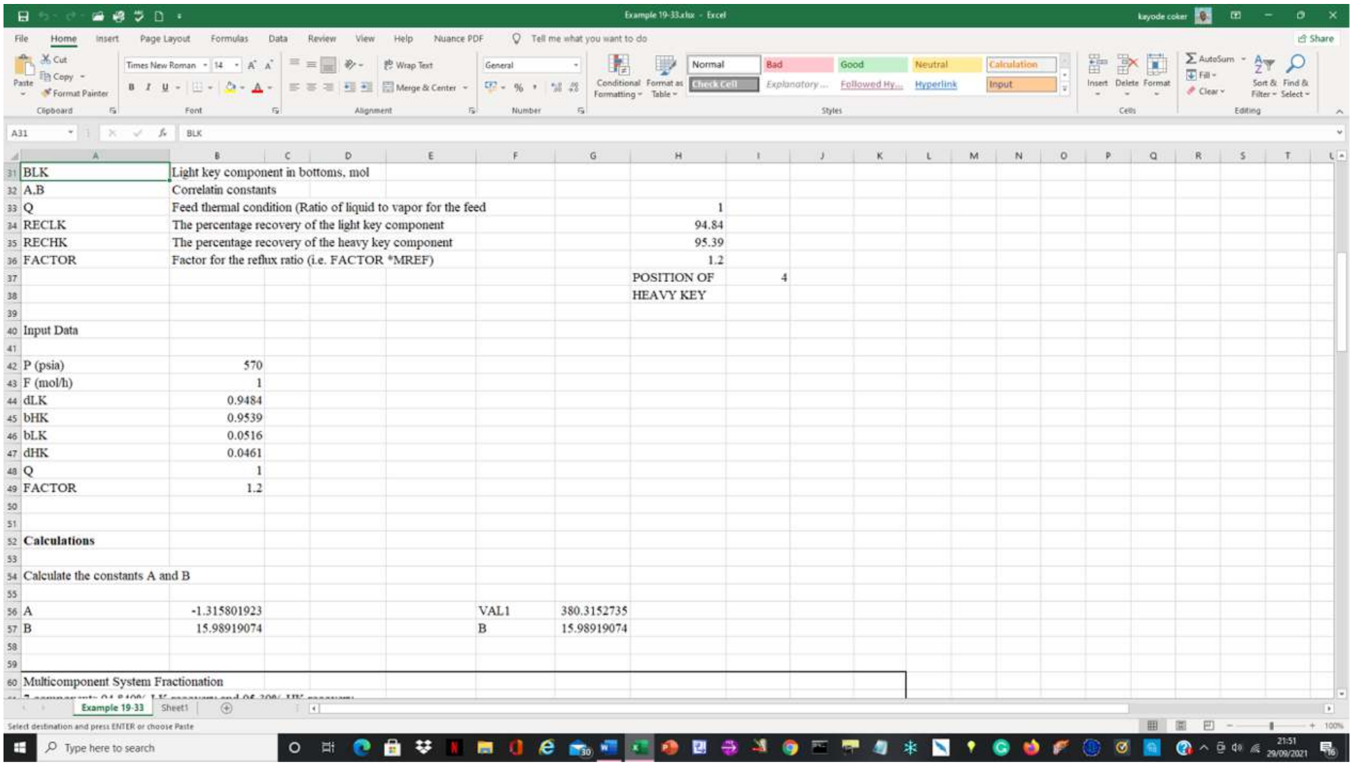


Rapid Estimates of Multi-Component Distillation and Minimum Number of Trays by A.K. Coker
The program will calculate the feed component recovery in both the column
distillate and bottoms products and estimate the minimum number of stages needed
to perform the given separation.
Feed streams containing up to 12 components can be handled

The following assumptions for these calculations are:
1. There is only one feed stream with 12 or fewer components.
2. There is only one heavy key component
3. Feed components are arranged in order of decreasing relative volatility, with
the light key (LK) and heavy key (HK) components shown adjacent to each other.
4. The feed enters the column at the optimal stage.
5. The column produces two products (distillate and bottoms) with overhead
condenser and bottom reboiler

NOTATION

ALPHAi	Relative volatility of component i
Ki	Equilibrium constant for component I in the feed stream
KHK	Equilibrium constant for heavy key component in the feed stream
Di	Component i in distillate, mol
Bi	Component i in bottoms, mol
BHK	Heavy key component in bottoms, mol
FHK	Heavy key component in the feed, mol
DLK	Light key component in the distillate, mol
FLK	Light key component in the feed, mol
ALPHA(LK)	Relative volatility of light key component
Fi	Component i in the feed
NUM1	Minimum trays at total reflux
DHK	Heavy key component in distillate, mol



BLK	Light key component in bottoms, mol	
A,B	Correlatin constants	
Q	Feed thermal condition (Ratio of liquid to vapor for the feed	1
RECLK	The percentage recovery of the light key component	94.84
RECHK	The percentage recovery of the heavy key component	95.39
FACTOR	Factor for the reflux ratio (i.e. FACTOR *MREF)	1.2

POSITION OF HEAVY KEY 4

Input Data

P (psia)	570
F (mol/h)	1
dLK	0.9484
bHK	0.9539
bLK	0.0516
dHK	0.0461
Q	1
FACTOR	1.2

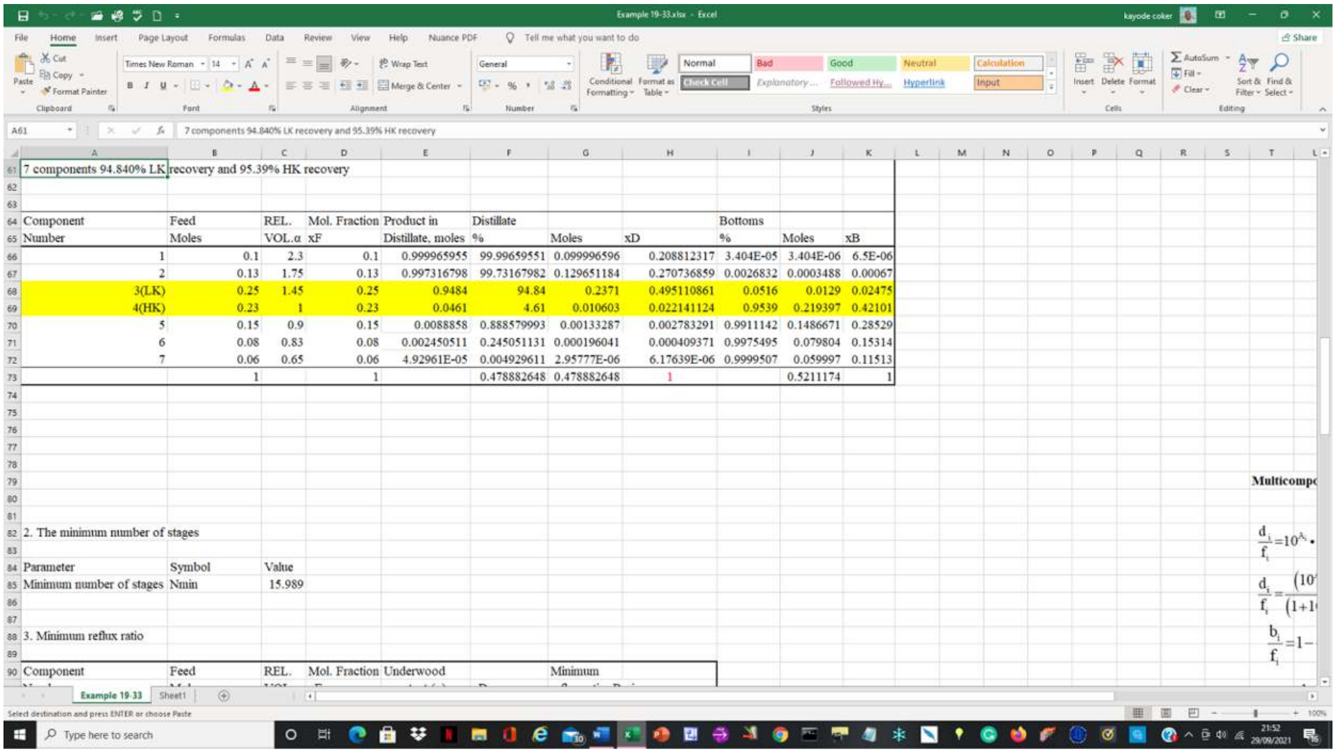
Calculations

Calculate the constants A and B

A	-1.315801923	VAL1	380.3152735
B	15.98919074	B	15.98919074

Multicomponent System Fractionation

Table 19.31 Excel spreadsheet calculations (Example 19.33.xlsx) of Example 19.33. (*Continued*)



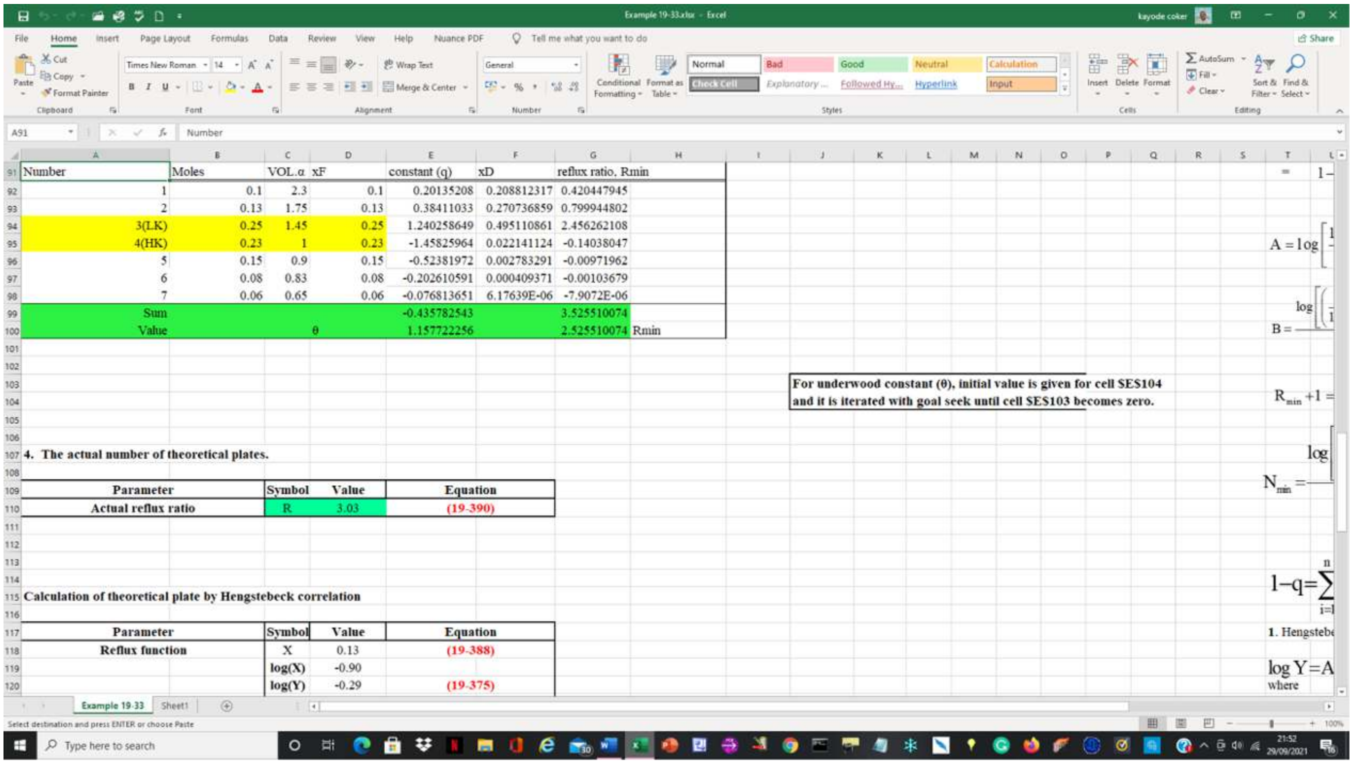
7 components 94.840% LK recovery and 95.39% HK recovery

Component Number	Feed Moles	REL. VOL.α	Mol. Fraction xF	Product in Distillate, moles	Distillate %	Moles	xD	Bottoms %	Moles	xB
1	0.1	2.3	0.1	0.999965955	99.99659551	0.099996596	0.208812317	3.404E-05	3.404E-06	6.5E-06
2	0.13	1.75	0.13	0.997316798	99.73167982	0.129651184	0.270736859	0.0026832	0.0003488	0.00067
3(LK)	0.25	1.45	0.25	0.9484	94.84	0.2371	0.495110861	0.0516	0.0129	0.02475
4(HK)	0.23	1	0.23	0.0461	4.61	0.010603	0.022141124	0.9539	0.219397	0.42101
5	0.15	0.9	0.15	0.0088858	0.888579993	0.00133287	0.002783291	0.9911142	0.1486671	0.28529
6	0.08	0.83	0.08	0.002450511	0.245051131	0.000196041	0.000409371	0.9975495	0.079804	0.15314
7	0.06	0.65	0.06	4.92961E-05	0.004929611	2.95777E-06	6.17639E-06	0.9999507	0.059997	0.11513
	1		1		0.478882648	0.478882648	1		0.5211174	1

2. The minimum number of stages

Parameter	Symbol	Value
Minimum number of stages	Nmin	15.989

3. Minimum reflux ratio



Component Number	Feed Moles	REL. VOL.α	Mol. Fraction xF	Underwood constant (q)	D xD	Minimum reflux ratio, Rmin	
1	0.1	2.3	0.1	0.20135208	0.208812317	0.420447945	
2	0.13	1.75	0.13	0.38411033	0.270736859	0.799944802	
3(LK)	0.25	1.45	0.25	1.240258649	0.495110861	2.456262108	
4(HK)	0.23	1	0.23	-1.45825964	0.022141124	-0.14038047	
5	0.15	0.9	0.15	-0.52381972	0.002783291	-0.00971962	
6	0.08	0.83	0.08	-0.202610591	0.000409371	-0.00103679	
7	0.06	0.65	0.06	-0.076813651	6.17639E-06	-7.9072E-06	
Sum				-0.435782543		3.525510074	
Value			θ	1.157722256		2.525510074	Rmin

For underwood constant (θ), initial value is given for cell E104 and it is iterated with goal seek until cell E103 becomes zero.

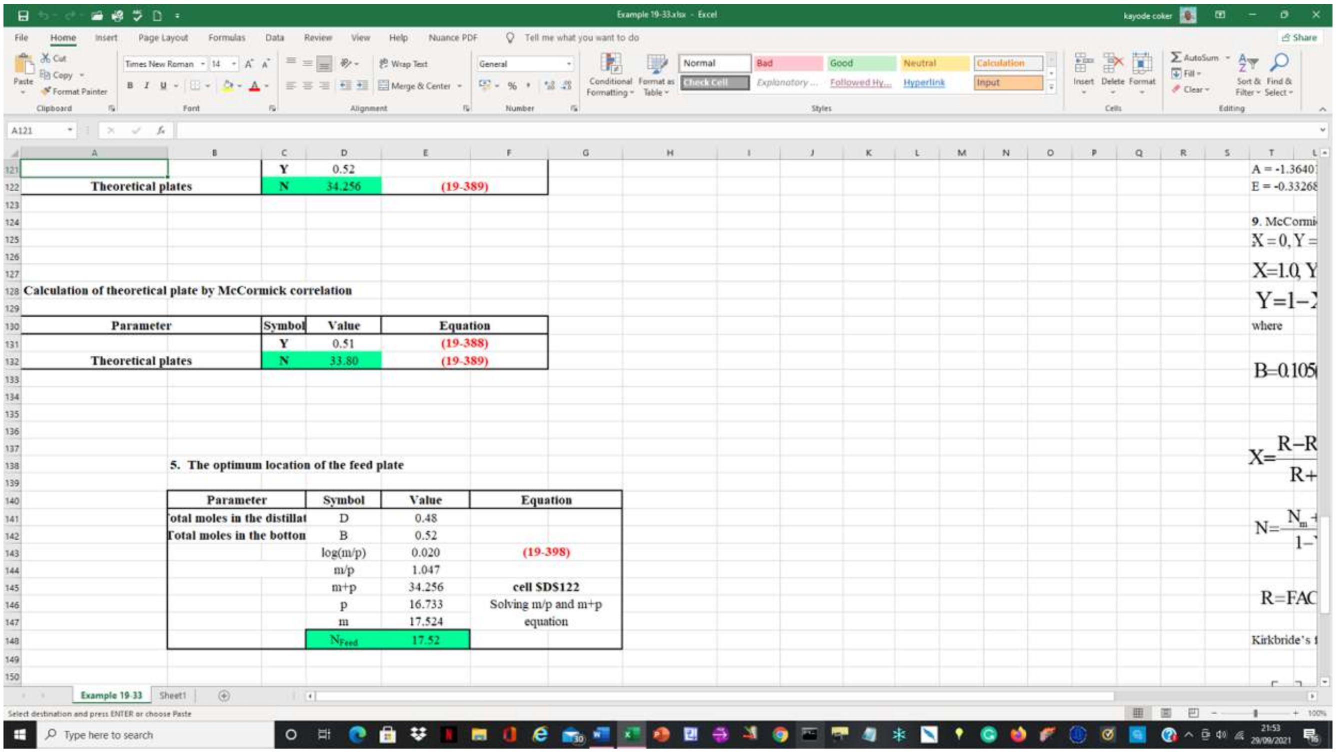
4. The actual number of theoretical plates.

Parameter	Symbol	Value	Equation
Actual reflux ratio	R	3.03	(19-390)

Calculation of theoretical plate by Hengstebeck correlation

Parameter	Symbol	Value	Equation
Reflux function	X	0.13	(19-388)
	log(X)	-0.90	
	log(Y)	-0.29	(19-375)

Table 19.31 Excel spreadsheet calculations (Example 19.33.xlsx) of Example 19.33. (*Continued*)



Parameter	Symbol	Value	Equation
	Y	0.52	
Theoretical plates	N	34.256	(19-389)

Calculation of theoretical plate by McCormick correlation

Parameter	Symbol	Value	Equation
	Y	0.51	(19-388)
Theoretical plates	N	33.80	(19-389)

5. The optimum location of the feed plate

Parameter	Symbol	Value	Equation
otal moles in the distillat	D	0.48	
Total moles in the botton	B	0.52	
	log(m/p)	0.020	(19-398)
	m/p	1.047	
	m+p	34.256	cell D122
	p	16.733	Solving m/p and m+p equation
	m	17.524	
	N_{Feed}	17.52	

A = -1.3640

E = -0.33268

9. McCormi

X = 0, Y =

X=1.0, Y

Y=1−

where

B=0.105

$X=\frac{R-R}{R+}$

$N=\frac{N_m+}{1-}$

R=FAC

Kirkbride's

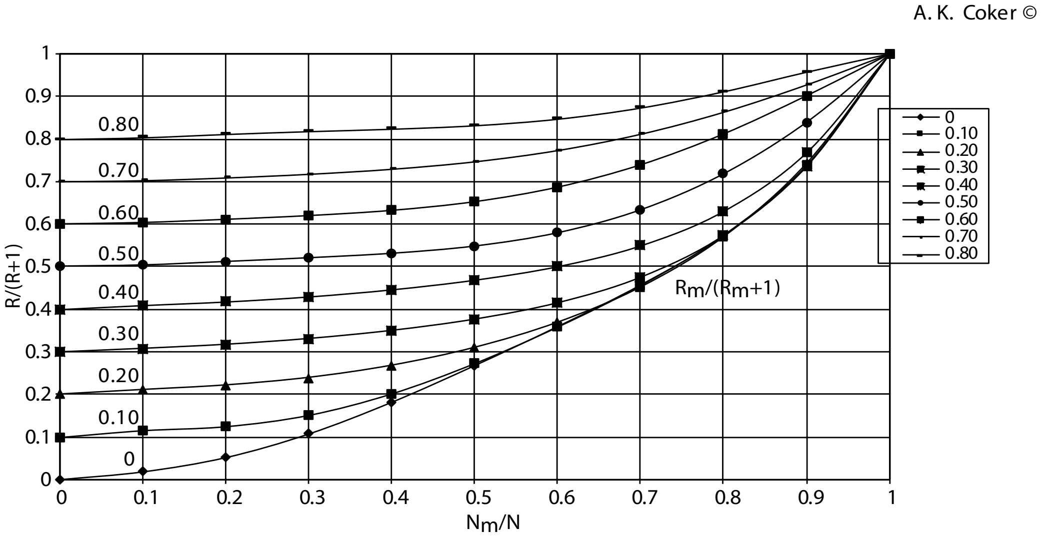


Figure 19.77 Plate–reflux correlation of Erbar and Maddox.

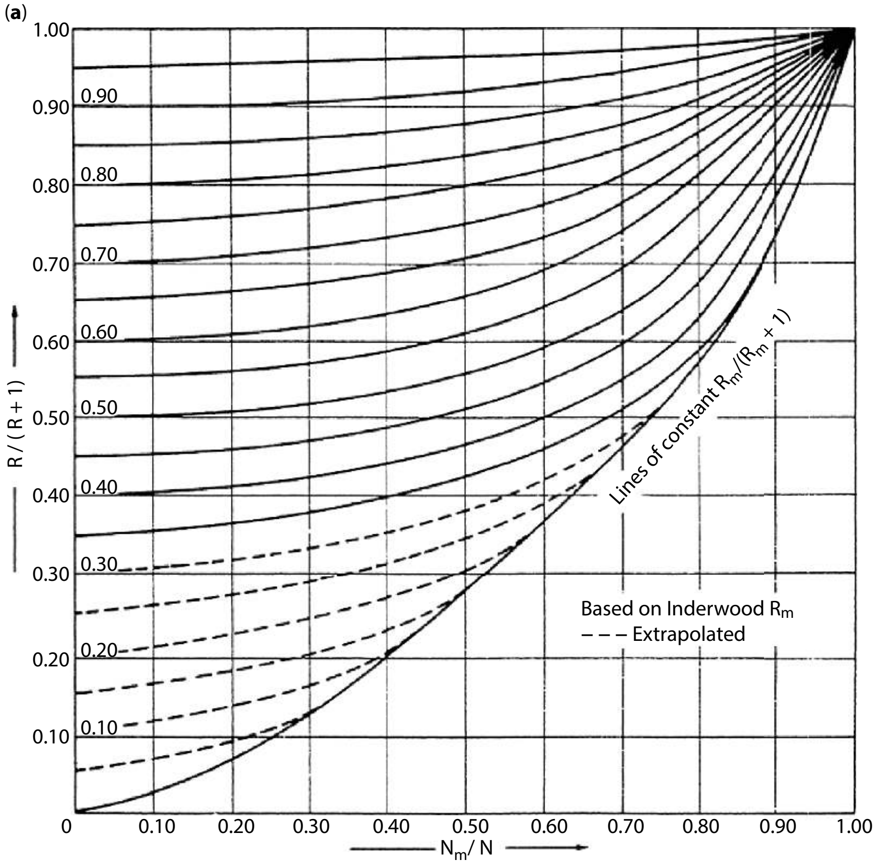


Figure 19.77a Erbar–Gray–Maddox correlation. Dashed lines: extrapolated (Source: Erbar and Maddox, Hydrocarbon Process, Petrol Refiner, 4, No. 5, 183, 1961).

and heat data for the components. This then provides an accurate analysis at each tray (liquid and vapor analysis) and also the heat duty of the bottoms reboiler and overhead total or partial condenser.

Torres-Marchal [155, 156] present a detailed graphical solution for multicomponent ternary systems that can be useful to establish the important parameters prior to undertaking a more rigorous solution with a computer program. This technique can be used for azeotropic mixtures, close-boiling mixtures, and similar situations.

An alternative improved solution for Underwood's method is given by Erbar, Joyner, and Maddox [157] with an example.

19.37.10 Erbar–Maddox Method [158]

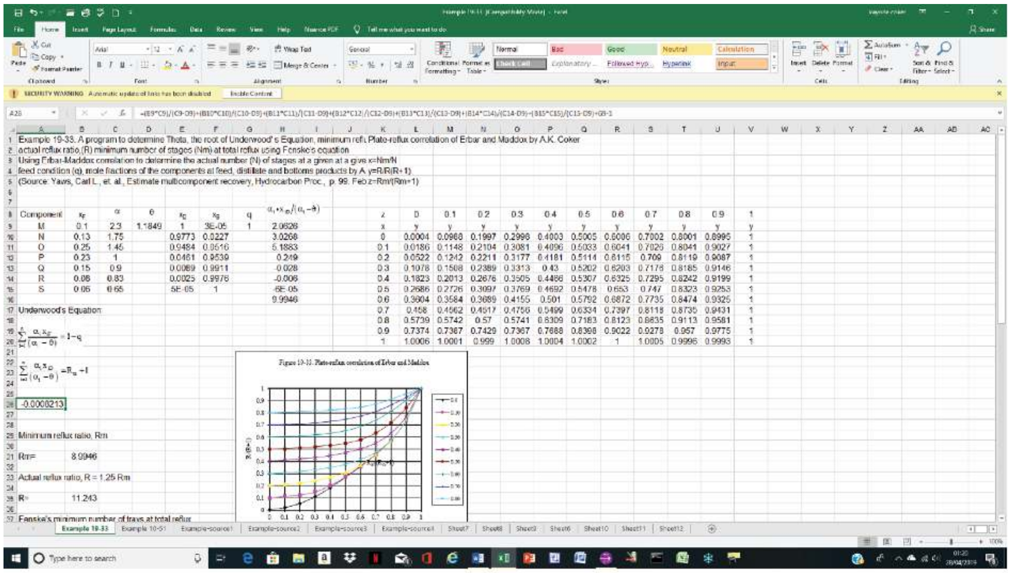
This method is used to determine the actual theoretical stages at a practical actual reflux ratio. It uses a plot of R/(R + 1) against N_{min}/N with $R_{min}/(R_{min}+1)$ as the parameter. When $R = R_{min}$, the axis becomes zero. Therefore, the y-axis of the diagram represents minimum reflux conditions. When $N = N_{min}$, both x and y coordinates become unity. Branan [159] has developed a polynomial expression that represents the Erbar–Maddox curves. He represents the equation by the following:

$$y = A + Bx + Cx^2 + Dx^3 + Ex^4 + Fx^5 \tag{19.402}$$

where

A, B, C, D, E, F = constants for varying values of z.

$$x = N_m/N \tag{19.403}$$



A28 =(B9*C9)/(C9-D9)+(B10*C10)/(C10-D9)+(B11*C11)/(C11-D9)+(B12*C12)/(C12-D9)+(B13*C13)/(C13-D9)+(B14*C14)/(C14-D9)+(B15*C15)/(C15-D9)+G9-1

1 Example 19-33. A program to determine Theta, the root of Underwood's Equation, minimum refl. Plate-reflux correlation of Erbar and Maddox by A.K. Coker
2 actual reflux ratio,(R) minimum number of stages (Nm) at total reflux using Fenske's equation
3 Using Erbar-Maddox correlation to determine the actual number (N) of stages at a given at a give x=Nm/N
4 feed condition (q), mole fractions of the components at feed, distillate and bottoms products by A. y=R/R(R+1)
5 (Source: Yaws, Carl L., et. al., Estimate multicomponent recovery, Hydrocarbon Proc., p. 99. Feb z=Rm/(Rm+1)

Component	x_F	α	θ	x_D	x_B	q	$\alpha_i x_{iD}/(\alpha_i-\theta)$
M	0.1	2.3	1.1849	1	3E-05	1	2.0626
N	0.13	1.75		0.9773	0.0227		3.0268
O	0.25	1.45		0.9484	0.0516		5.1883
P	0.23	1		0.0461	0.9539		0.249
Q	0.15	0.9		0.0089	0.9911		-0.028
R	0.08	0.83		0.0025	0.9976		-0.006
S	0.06	0.65		5E-05	1		-6E-05
							9.9946

z	0	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9	1
x	y	y	y	y	y	y	y	y	y	y	y
0	0.0004	0.0968	0.1997	0.2996	0.4003	0.5005	0.6006	0.7002	0.8001	0.8995	1
0.1	0.0186	0.1148	0.2104	0.3081	0.4096	0.5033	0.6041	0.7026	0.8041	0.9027	1
0.2	0.0522	0.1242	0.2211	0.3177	0.4181	0.5114	0.6115	0.709	0.8119	0.9087	1
0.3	0.1078	0.1508	0.2389	0.3313	0.43	0.5202	0.6203	0.7176	0.8185	0.9146	1
0.4	0.1823	0.2013	0.2676	0.3505	0.4486	0.5307	0.6325	0.7295	0.8242	0.9199	1
0.5	0.2686	0.2726	0.3097	0.3769	0.4692	0.5478	0.653	0.747	0.8323	0.9253	1
0.6	0.3604	0.3584	0.3689	0.4155	0.501	0.5792	0.6872	0.7735	0.8474	0.9325	1
0.7	0.458	0.4562	0.4517	0.4756	0.5499	0.6334	0.7397	0.8118	0.8735	0.9431	1
0.8	0.5739	0.5742	0.57	0.5741	0.6309	0.7183	0.8123	0.8635	0.9113	0.9581	1
0.9	0.7374	0.7367	0.7429	0.7367	0.7688	0.8398	0.9022	0.9278	0.957	0.9775	1
1	1.0006	1.0001	0.999	1.0008	1.0004	1.0002	1	1.0005	0.9996	0.9993	1

17 Underwood's Equation

$$\sum_{i=1}^{n}\frac{\alpha_i x_{iF}}{(\alpha_i-\theta)}=1-q$$

$$\sum_{i=1}^{n}\frac{\alpha_i x_{iD}}{(\alpha_i-\theta)}=R_m-1$$

26 -0.0000213

29 Minimum reflux ratio, Rm

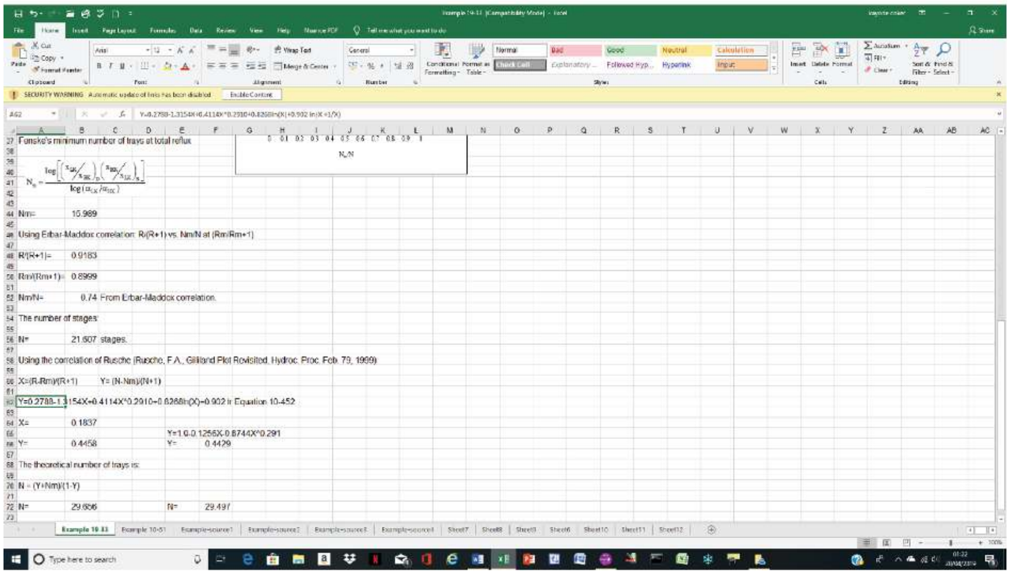
31 Rm= 8.9946

33 Actual reflux ratio, R = 1.25 Rm

35 R= 11.243

37 Fenske's minimum number of trays at total reflux

(**a**)



A62 Y=0.2788-1.3154X+0.4114X^0.2910+0.8268ln(X)+0.902 ln(X+1/X)

37 Fenske's minimum number of trays at total reflux

$$N_m=\frac{\log\left[\left(x_{LK}/x_{HK}\right)_D\left(x_{HK}/x_{LK}\right)_B\right]}{\log(\alpha_{LK}/\alpha_{HK})}$$

44 Nm= 15.989

46 Using Erbar-Maddox correlation: R/(R+1) vs. Nm/N at (Rm/Rm+1)

48 R/(R+1)= 0.9183

50 Rm/(Rm+1)= 0.8999

52 Nm/N= 0.74 From Erbar-Maddox correlation.

54 The number of stages:

56 N= 21.607 stages.

58 Using the correlation of Rusche (Rusche, F.A., Gilliland Plot Revisited, Hydroc. Proc. Feb. 79, 1999)

60 X=(R-Rm)/(R+1) Y= (N-Nm)/(N+1)

62 Y=0.2788-1.3154X+0.4114X^0.2910+0.8268ln(X)+0.902 ln Equation 10-452

64 X= 0.1837

65 Y=1.0-0.1256X-0.8744X^0.291

66 Y= 0.4458 Y= 0.4429

68 The theoretical number of trays is:

70 N = (Y+Nm)/(1-Y)

72 N= 29.656 N= 29.497

(**b**)

Figure 19.78 (a) Excel program snapshot of Example 19.33. (b) Excel program snapshot of Example 19.33.

$$y = R/(R+1) \tag{19.404}$$

$$z = R_m/(R_m+1) \tag{19.405}$$

Table 19.32 shows the constant values at varying z. The author has used the values in Table 19.32 and developed an Excel spreadsheet to create the Erbar–Maddox correlation. Figure 19.77 shows the plates–reflux correlation of Erbar–Maddox and is based on a bubble-point feed. The correlations from Figure 19.77 are in good agreement with the Erbar–Maddox correlation of Figure 19.77a. For other types of feed, the following correlation is used [130].

$$V_u = V_k + \frac{(1-D/F)(H_{F,u}-H_{F,k})}{Q_c/L_o} \tag{19.406}$$

Table 19.32 Constants for the Erbar-Maddox correlation.

z	A	B	C	D	E	F
0	0.00035	0.16287	-0.23193	5.09032	-8.50815	4.48718
0.1	0.09881	0.32725	-2.57575	10.20104	-12.8205	5.76923
0.2	0.19970	0.14236	-0.58646	2.60561	-3.12499	1.76282
0.3	0.29984	0.09393	-0.23913	1.49008	-2.4388	1.79486
0.4	0.40026	0.12494	-0.49585	2.15836	-3.27068	2.08333
0.5	0.50049	-0.03058	0.81585	-2.61655	3.61305	-1.28205
0.6	0.60063	-0.00792	0.60063	-2.06912	3.39816	-1.52243
0.7	0.70023	-0.01109	0.45388	-1.25263	1.94348	-0.83334
0.8	0.80013	-0.01248	0.76154	-2.72399	3.85707	-1.68269
0.9	0.89947	0.00420	0.38713	-1.14962	1.40297	-0.54487
1.0	1.0	0	0	0	0	0

Source: Branan, C.R., Rules of Thumb for Chemical Engineers: A manual of quick, accurate solutions to everyday engineering problems, 4th Ed., Elsevier, 2005.

where

D = Distillate flow rate, lb-mole/h
F = Feed flow rate, lb-mole/h
H = Enthalpy, Btu/h
L_o = Liquid condensed at the condenser, lb-mole/h
Q_c = Condenser duty, Btu/h
V = Vapor flow rate in the rectifying section, lb-mole/h

Subscripts

F = feed.
k = at the bubble point.
u = at the conditions differing from bubble point.

In the Erbar–Maddox correlation, minimum stages are calculated by the Winn method and minimum reflux by the Underwood method. But the Fenske minimum stages method can also be used.

19.37.11 Underwood Algebraic Method: Adjacent Key Systems [144]

This system for evaluating multicomponent adjacent key systems, assuming constant relative volatility and constant molal overflow, has proven generally satisfactory for many chemical and hydrocarbon applications. It gives a rigorous solution for constant molal overflow and volatility, and acceptable results for most cases, which deviate from these limitations.

Overall Column–Constant α

$$(L/D)_{min} + 1 = \frac{(\alpha_a x_a)_D}{\alpha_a - \theta} + \frac{(\alpha_b x_b)_D}{\alpha_b - \theta} + \ldots \frac{(\alpha_i x_i)_D}{\alpha_i - \theta} \tag{19.407}$$

In arriving at $(L/D)_{min}$, the correct value of θ is obtained from

$$1-q=\frac{(\alpha_a x_a)_F}{\alpha_a-\theta}+\frac{(\alpha_b x_b)_F}{\alpha_b-\theta}+\ldots\frac{(\alpha_i x_i)_F}{\alpha_i-\theta}=\sum\frac{x_{Fi}}{1-\theta/\alpha_i} \tag{19.408}$$

The "q" value is previously described for the thermal condition of the feed.

Rectifying section only:

$$V_r=\sum_{i=l,h,H}\frac{Dx_{Di}}{1-\theta/\alpha_i} \tag{19.409}$$

Stripping section only:

$$V_s=\sum_{i=l,h,H}\frac{Bx_{Bi}}{1-\theta/\alpha_i} \tag{19.410}$$

At the minimum reflux condition, all the θ values are equal and generally related:

$$\alpha_h < \theta < \alpha_l$$

Suggested Procedure

1. From Equation (19.407), expressing θ and q evaluate θ by trial and error, noting that θ will have a value between the α of the heavy key and the α of the light key evaluated at or near pinch temperatures, or at α_{avg}. Suggested tabulation, starting with an assumed θ value, θ_a:

Component	$\mathbf{x_{Fi}}$	$\boldsymbol{\alpha_i x_{Fi}}$	$\boldsymbol{\alpha_i - \theta}$	$\boldsymbol{\alpha_i x_{Fi}/(\alpha_i-\theta)}$	$\boldsymbol{\alpha_i x_{Fi}/(\alpha_i-\theta)^2}$
a	x_{Fa}	$\alpha_a x_{Fa}$	$\alpha_a-\theta$	$\alpha_a x_{Fa}/(\alpha_a-\theta)$	$\alpha_a x_{Fa}/(\alpha_a-\theta_a)^2$
b	x_{Fb}	$\alpha_b x_{Fb}$	$\alpha_b-\theta$	$\alpha_b x_{Fb}/(\alpha_b-\theta)$	$\alpha_b x_{Fb}/(\alpha_b-\theta_b)^2$
•	•	•	•	•	•
•	•	•	•	•	•
				$\sum\Psi(\theta_a)$	$\sum\Psi'(\theta_a)$

Ψ, Ψ′ represent function.
Corrected θ by Newton's approximation method:

$$\theta_c = \theta(\text{assumed}) - \frac{\psi(\theta_a)}{\psi'(\theta_a)} \tag{19.411}$$

Repeat the same type of tabular computation, substituting the corrected θ_c for the θ_a. If the second corrected θ_c' checks closely with θ_c the right value of θ has been obtained; if not, a third recalculation should be made using the θ_c' value as the new assumed value.

Note that average α values should be used (constant) for each component unless the values vary considerably through the column. In this latter case, follow the discussion given elsewhere in this section.

2. Calculate $(L/D)_{min}$ by substituting the final θ value in Equation (19.407). Note that this requires evaluating the functions associated with θ at the composition of the distillate product. The α values are the constant values previously used above.

19.37.12 Underwood Algebraic Method: Adjacent Key Systems; Variable α

For varying α systems, the following procedure is suggested:

1. Assume $(L/D)_{min}$ and determine the pinch temperature by Colburn's method.
2. At this temperature, evaluate α at pinch and α at overhead temperature, obtaining a geometric average α. As an alternative, Shiras *et al*. [160] indicate a t_{avg} value which gives acceptable results when compared to pinch and stepwise calculations. This suggestion calculates

$$t_{avg} = (D_{t_o} + B_{t_B})/F$$

3. Determine Underwood's θ value as previously described using the average α value.
4. Calculate $(L/D)_{min}$ and compare with assumed value of (1) above. If check is satisfactory, $(L/D)_{min}$ is complete; if not, reassume a new $(L/D)_{min}$ using calculated value as basis, and repeat (1) through (4) until satisfactory check is obtained.

where

t_o = overhead temp °F
t_B = bottoms temp °F
t_{avg} = avg temp, °F

To aid in solving the tedious Underwood equation to ultimately arrive at (L/D)min, Frank [84] has developed Figure 19.79, which applies for liquid feed at its bubble point and to binary or multicomponent systems, but does require that the key components are adjacent. Otherwise, the system must be solved for two values of θ [82]. To obtain the necessary parameters for Figure 19.79, Frank recommends using the same overhead concentrations that were used in or calculated by the Fenske equation for the Underwood solution (θ = Underwood constant).

Example 19.34

An equimolar binary mixture with a relative volatility of 2 is fed to a distillation column, which produces a distillate containing 97.5% of the more volatile component and a residue containing 10% of the more volatile component. The feed is liquid at its boiling point, and 80% of the liquid entering the reboiler is vaporized and returned to the column as boil-up.

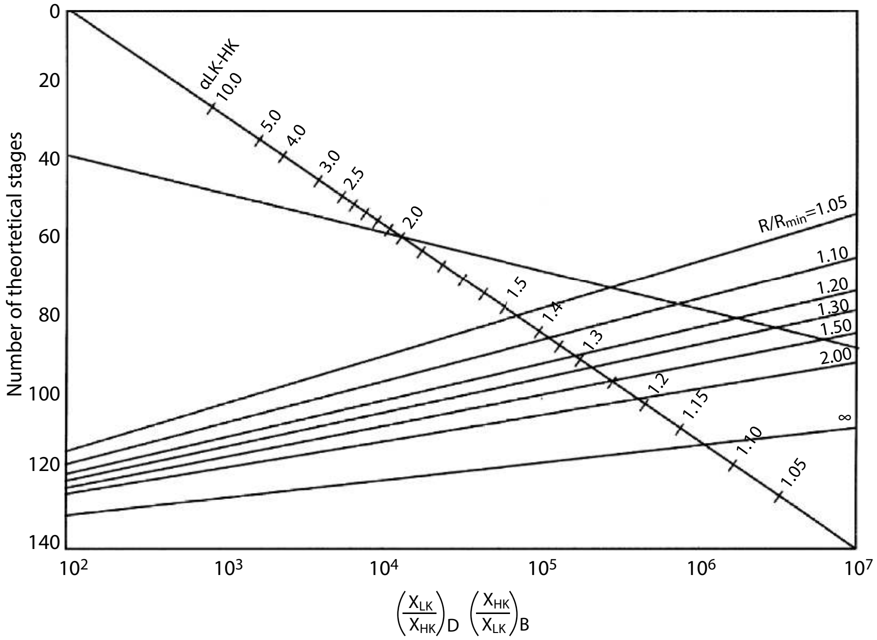
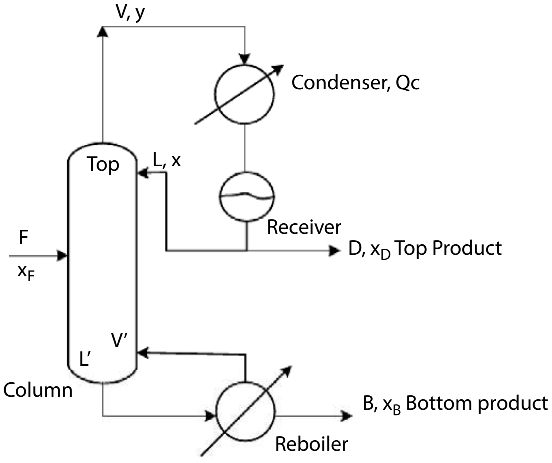


Figure 19.79 Short-cut solution of Fenske–Underwood–Gilliland theoretical trays for multicomponent distillation. (Used by permission, Frank, O., Chem. Eng., Mar. 14, p. 109 (1977)).

(a) Estimate the number of theoretical plates provided by this column.

(b) Check your answer in part (a) by reference to the graph of the Gilliland correlation and Erbar and Maddox correlation.

Solution



(a) The overall mass balance on the column.

$$F = D + B \tag{1}$$

The component balance on the more volatile component (MVC) is

$$Fx_F = Dx_D + Bx_B \tag{2}$$

Assuming that F = 100 mol/h.

$$x_F = 0.5,\ x_D = 0.975,\text{ and } x_B = 0.1$$

Substituting the values of the components in Equation (2) gives

$$100(0.5) = D(0.975) + B(0.1)$$

$$50 = 0.975D + 0.1B \tag{3}$$

Rearranging Equation (1) and substituting into Equation (3) gives

$$50 = 0.975D + 0.1(100\text{-}D) \tag{4}$$

$$50 = 0.975D + 10\text{-}0.1D \tag{5}$$

$$D = 45.71\text{ mol/h.}$$

$$B = 54.29\text{ mol/h.}$$

The upper operating line (UOL) at the top of the column is

The overall mass balance on UOL is

$$V = L + D \tag{6}$$

Component balance is

$$V_n y_n = L_{n+1} x_{n+1} + D x_D \tag{7}$$

or

$$y_n = \frac{L_{n+1}}{V_n} x_{n+1} + \frac{D}{V_n} x_D \tag{8}$$

$$\text{Reflux ratio, } R = \frac{L}{D} \tag{9}$$

Substituting Equation (9) into Equation (6) gives

$$V_n = RD + D \tag{10}$$

and Equation (8) (UOL), assuming constant molar overflow, becomes

$$y_n = \frac{R}{(R+1)} x_{n+1} + \frac{x_D}{R+1} \tag{11}$$

The intercept on the UOL is

$$I = \frac{x_D}{(R+1)} \tag{12}$$

Since the feed is liquid at its boiling point (i.e., q = 1) and 80% of the liquid entering the reboiler (L′) is vaporized to give the vapor (V′), which leaves the reboiler and returns to the column.

Therefore, at the bottom of the column, if L′ = 1.0 and V′ = 0.8.

The mass balance at the bottom of the column with the reboiler is

$$L' = V' + B \tag{13}$$

or

$$V' = L' - B \tag{14}$$

The component balance at the bottom is

$$V'y_m = L'x_{m+1} - Bx_B \tag{15}$$

$$y_m = \frac{L'}{V'}x_{m+1} - \frac{B}{V'}x_B \tag{16}$$

$$\text{Slope is } \frac{L'}{V'} = \frac{1}{0.8} = 1.25$$

where L′/V′ = gradient of lower operating line (LOL). The McCabe-Thiele diagram of Figure 10.96 is drawn as follows:

At the coordinates (x_w, x_w) = 0.1 on the 45° line, draw the LOL with the slope of 1.25, to intercept the "q" (i.e., q = 1 at boiling point) line at $x_F = 0.5$.

At coordinates (x_D, x_D) = 0.975 on the 45° line, draw the UOL to intersect the LOL at q =1 and extend to the intercept I on the y-ordinate of the VLE diagram.

Step-off the line between the UOL, LOL and the 45° line from $x_D = 0.975$ to $x_W = 0.1$ to determine the number of equilibrium stages.

$$I = \frac{x_D}{R+1} = 0.22 \text{ (from McCabe-Theile diagram)}$$

$$\frac{0.975}{(R+1)} = 0.22$$

$$R = 3.43$$

From McCabe-Thiele diagram (Figure 19.80):

$$N + 1 = 14 \text{ stages}$$

$$N = 13 \text{ stages.}$$

(b) R_{min}: from the points (x_D, x_D), a straight line is drawn that touches the VLE line with the "q" line and extended to the y-axis at a point $I_{min} = \frac{x_D}{(R_m + 1)}$.

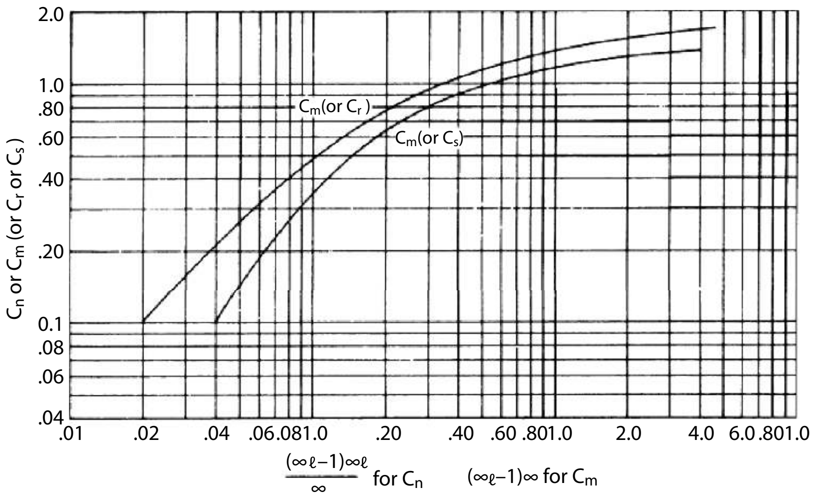


Figure 19.80 Colburn minimum reflux factors, above (C_n) and (C_m) feed point. (Used by permission, Colburn, A. P., The American Institute of Chemical Engineers, Trans., Amer. Inst. Chem. Eng., Vol. 37, p. 805, (1941). All rights reserved).

$$I_{min} = \frac{x_D}{(R_m + 1)} = 0.34$$

$$\frac{0.975}{(R_m + 1)} = 0.34$$

$$R_m = 1.86$$

Using the Underwood equation for the minimum reflux ratio:

$$\begin{aligned} R_m &= \frac{1}{(\alpha - 1)}\left[\frac{x_{LK,D}}{x_{LK,F}} - \frac{\alpha(1 - x_{LK,D})}{(1 - x_{LK,F})}\right] \\ &= \frac{1}{(2-1)}\left[\frac{0.975}{0.5} - \frac{2(1-0.975)}{(1-0.5)}\right] \\ R_m &= 1.85 \end{aligned} \tag{19.210}$$

The Fenske method for the minimum number of theoretical stages at total reflux is

$$\begin{aligned} N_m &= \frac{\ln\left[(x_{LK}/x_{HK})_D (x_{HK}/x_{LK})_B\right]}{\ln(\alpha_{LK}/\alpha_{HK})_{avg}} \\ &= \frac{\ln\left[\left(\frac{0.975}{0.025}\right)\left(\frac{0.9}{0.1}\right)\right]}{\ln 2} \\ N_m &= 8.5 \end{aligned} \tag{19.182}$$

From the Gilliland correlation (Figure 19.76), the abscissa axis is

$$\frac{R - R_m}{R + 1} = \frac{3.66 - 1.85}{(3.66 + 1)} = 0.388 \tag{19.219}$$

The corresponding point on the ordinate axis of the Gilliland correlation is 0.31.

Therefore, the number of theoretical stages can be determined from the ordinate by

$$\begin{aligned}&\frac{N-N_m}{N+1}=0.31\\&\frac{N-8.45}{N+1}=0.31\\&N=12.69\,(13\text{ stages})\end{aligned}\tag{19.220}$$

Using the Erbar and Maddox correlation (Figure 19.77 or 19.77a)

The lines of constant $\frac{R_m}{R_m+1}$ is

1.85/2.85 = 0.649 (0.65)

The ordinate axis of Erbar and Maddox correlation is

$$\frac{R}{R+1}=\frac{3.66}{(3.66+1)}=0.79$$

A line is drawn at point 0.79 on the ordinate of the correlation and intercepts the line of constant $\frac{R_m}{R_m+1}$ at 0.65. The corresponding point on the abscissa is 0.74.

The number of theoretical stages is

$$\frac{N_m}{N}=0.74$$

N = 8.45/0.74 = 11.4 stages.

Example 19.35: Using the Data from Example 19.20 (Erbar–Maddox correlation Figure 19.77a)

Using the Erbar and Maddox correlation.

The lines of constant $\frac{R_m}{R_m+1}$ is

1.85/2.85 = 0.649 (0.65)

The ordinate axis of Erbar and Maddox correlation is

$$\frac{R}{R+1}=\frac{3.66}{(3.66+1)}=0.79$$

A line is drawn at point 0.79 on the ordinate of the correlation and intercepts the line of constant $\frac{R_m}{R_m+1}$ at 0.65. The corresponding point on the abscissa is 0.74. The number of theoretical stages is

$$\frac{N_m}{N}=0.74$$

$$N=8.45/0.74=11.4\text{ stages.}$$

19.37.13 Underwood Algebraic Method: Split Key Systems: Constant Volatility [144]

This method appears tedious, but it is not so unwieldy as to be impractical, but it does require close attention to detail. However, a value of $(L/D)_{min}$ can be obtained with one trial that may be satisfactory for "order of magnitude" use, which is often all that is needed before proceeding with detailed column design and establishment of operational L/D ratio.

1. Assume θ_f values and check by

$$\sum \frac{\alpha_i x_{Fi}}{\alpha_i - \theta_{fi}} = 1 - q \tag{19.412}$$

There are total solutions of θ_{fi} equal to one more than the number of split components between the keys. The θ_f values will be spaced:

$$\alpha_{13}\ \theta_{f3}\ \alpha_4\ \theta_{f4}\ \alpha_5\ \theta_{f5}\ \alpha_{h6}$$

where α_l is the light key and component number 3, and correspondingly for the heavy key, component number 6. Determine θ values as for constant volatility case of adjacent keys.

For some systems, the θ values can be assumed without further solution of the above relation, but using these assumed values as below.

2. Calculate

$$v = \frac{\dfrac{\dfrac{1}{(P)(\theta_{fi})}}{i = h - 1}}{\dfrac{\dfrac{1+1}{(P)(\alpha_i)}}{i = h - 1}} \tag{19.413}$$

which represents (for the hypothetical system set up in (1)) the product (θ_{f5}) (θ_{f4}) (θ_{f3}) divided by the product of (α_5) (α_4), based upon the lightest component being numbered one, the next two, etc., the heaviest components having the higher numbered subscripts. P means product, and 1, i = h − 1, i = 1 + 1 are limits for evaluation referring to components between the keys, and the light and heavy keys.

3. Calculate

$$\omega_j = \frac{\dfrac{\dfrac{1+1}{(P)}}{i = h - 1}\left(1 - \dfrac{\alpha_i}{\alpha_j}\right)}{\dfrac{\dfrac{1}{(P)}}{i = h - 1}\left(1 - \dfrac{\theta_{fi}}{\theta_j}\right)} \tag{19.414}$$

For the θ example shown in (1) above:

$$\omega_3(\text{light key}) = \frac{(1-\alpha_5/\alpha_3)(1-\alpha_4/\alpha_3)}{(1-\theta_{f5}/\alpha_3)(1-\theta_{f4}/\alpha_3)(1-\theta_{f3}/\alpha_3)}$$

Also calculate ω for all components lighter than light key.

$$\omega_3(\text{heavy key}) = \frac{(1-\alpha_5/\alpha_6)(1-\alpha_4/\alpha_6)}{(1-\theta_{f5}/\alpha_6)(1-\theta_{f4}/\alpha_6)(1-\theta_{f3}/\alpha_6)}$$

Component, j	ω_j	α_j	$\frac{\omega_j}{\alpha_j} x_{D_j}$	$\omega j x_{Dj}$	$\frac{\omega_j}{\alpha_j} x_{D_j}$
l (light key)	•	•	•	•	•
h (heavy key)	•	•	•	•	•
L_{l+1} light					
L_{l+2} than light	•	•	•	•	•
L_{l+3} key, etc.				$\sum \omega_j x_{Dj}$	$\sum \frac{\omega_j}{\alpha_j}(x_{Dj})$

4. Calculate $(L/V)_{min}$: (internal)

$$(L/V)_{min} = (v)\frac{\sum\limits_{j=h,l,L} D\left(\frac{\omega_j}{\alpha_j}\right)\left(x_{D_j}\right)}{\sum\limits_{j=h,l,L} D(\omega_j)\left(x_{D_j}\right)} \tag{19.415}$$

5. Calculate External (L/D)min:

$$(L/D)_{min} = \frac{1}{(V/L)_{min} - 1} \tag{19.416}$$

For variable α conditions, the pinch temperature can be used for α determinations as previously described.

Example 19.36: Minimum Reflux Ratio Using Underwood Equation: Calculate the Minimum Reflux Ratio

Use $\phi_{fl} = 0.584$ to begin (assumed).

Expanding to determine more exact value of ϕ_{fl}

$$\Omega\phi = \Omega\phi_a + \phi - \phi_a \Omega'\phi_a$$

$$\Omega\phi = \sum \frac{\alpha_i x_f}{(\alpha_i - \phi)} - (1 - q)$$

$$\phi_{fl} = \frac{\phi_a - \Omega\phi_a}{\Omega'\phi_a}$$

$$\Omega'\phi = \sum \frac{\alpha_i x_f}{(\alpha_i - \phi)^2} \tag{19.417}$$

Component	X_{fi}	$\alpha_i X_{fi}$	$\alpha_i - \phi_a$	$\left[\frac{\alpha_i X_{fi}}{\alpha_i - \phi_a}\right]$	$\left[\frac{\alpha_i X_{fi}}{(\alpha_i - \phi_a)^2}\right]$
1	0.10	0.025	−0.334	−0.0749	+0.224
2	0.225	0.1125	−0.084	−1.34	+15.9
3	0.450	0.450	+0.416	+1.08	+2.6
4	0.225	0.450	+1.416	+0.318	+0.224
				Σ = −0.016	+18.948 = $\Omega'\phi_a$
			$\Omega\phi_a$ = −0.016 − (1 − q)		
			ϕ_a = −0.016 − (1 − 1)		
			ϕ_a = −0.016		

$$\phi_{fl\text{-corrected}} = 0.584 - (-0.016)/(18.948) - 0.584 + 0.00084 = 0.58484$$

Now use this new value of ϕ_{fl} in Underwood's equations;

$$\sum_{i=h,l,L} \frac{Dx_{Di}}{1 - \frac{\phi_r}{\alpha_i}} - V_r = 0 \tag{19.418}$$

For minimum reflux: $\phi_{fl} = \phi_r = \phi_s$

From calculations of related problem* the value of Dx_{Di} has been calculated:

Dx_{Di} = 0.01072 for heavy key
Dx_{Di} = 0.428 for light key
Dx_{Di} = 0.225 for lighter than light key.

$$\left[\frac{0.01072}{1 - \frac{0.5848}{0.50}} + \frac{0.428}{1 - \frac{0.5848}{1.0}} + \frac{0.225}{1 - \frac{0.5848}{2.0}}\right] - V_r = 0$$

$$[-0.638 + 1.03 + 0.318] - V_r = 0$$

From related problem, D has been determined to be = 0.6657 mol/mol feed.

$$-V_r = -1.285$$
$$V_r = 1.285$$
$$(V_r)_{min} = (L_r)_{min} + D$$
$$(L_r)_{min} = 1.285 - 0.6637 = 0.6213$$
$$\left(\frac{L_r}{D}\right)_{min} = \frac{0.6213}{0.6637} = 0.94, \text{ Minimum Reflux Ratio}$$

19.37.14 Minimum Reflux Colburn Method: Pinch Temperatures [161]

This method has also found wide usage and might be considered less tedious by some designers. It also yields an approximation of the rectifying and stripping section pinch temperatures. For adjacent keys,

Rectifying:

$$\left(\frac{L}{D}\right)_{min} = \frac{1}{\alpha - 1}\left(\frac{x_D}{x_n} - \alpha\frac{x_{hD}}{x_{hn}}\right) \quad (19.419)$$

where

α = relative volatility of any component referenced to the heavy key component
x_{hD} = overhead composition of heavy key component, mol fraction
x_{hn} = pinch composition of heavy key component, mol fraction
x_D = overhead composition of any light component, mol fraction
x_n = pinch composition of any light component, mol fraction

1. Calculate D, B, Dx_{Di}, and Bx_{Bi} from problem specification.
2. Assume or set the operating pressure and overhead temperature (may be calculated).
3. Calculate the liquid and vapor quantities and their respective compositions in the feed to the column.
4. Calculate estimated ratio of key components on feed plate based on the liquid portion of the feed.

$$r_f = \frac{\text{Mol fraction light key}}{\text{Mol fraction heavy key}}$$

   a. For all liquid feed at feed tray temperature (boiling point)
      r_f = mol fraction ratios in feed.
   b. For a part or all vapor feed just at its dew point,
      r_f = ratio of key components in the equilibrium liquid phase of feed.
   c. For all liquid feed below feed plate temperature
      r_f = ratio of key components at intersection point of operating line (from a McCabe–Thiele diagram).
5. Determine approximate pinch zone liquid composition for light key component

$$x_{ln} = \frac{r_f}{(1 + r_f)\left(1 + \sum \alpha_i x_{Fi}\right)_H} \quad (19.420)$$

$$\sum \alpha_i x_{fi} = \text{sum of } (\alpha_h \times \alpha_i x_{fi}) + (\alpha_{h+2} \times \alpha_i x_{fi}) + \ldots$$

for all components in liquid portion of feed heavier than heavy key. Note that x_{Fi} values are the mol fractions of the component in the liquid portion of feed only and the Σx_{Fi} is equal to 1.0.

6. Calculate approximate value for $(L/D)_{min}$.

$$(L/D)_{min} = \frac{1}{\alpha_i - 1}\left(\frac{x_{lD}}{x_{ln}} - \alpha\frac{x_{hD}}{x_{hn}}\right)$$

The second term in the right-hand parentheses can be omitted unless the mol fraction of the heavy key in the distillate, x_{hD}, is 0.1 or greater. Use $x_{hn} = x_{ln}/r_f$.

7. Estimate stripping and rectifying pinch temperatures at values one-third and two-thirds of the interval between the column bottoms and overhead, respectively.

8. Calculate internal vapor and liquid flows.

$$(L_r/D)_{min} = \text{assumed}$$

$$\text{Solve for } (L_r/V_r)_{min}$$

$$(L_r/V_r)_{min}) = \frac{1}{1 + (D/L_r)_{min}} \tag{19.421}$$

$$L_r = (\text{number})\ (V_r) = (L_r/D)_{min}(D)$$

D is known
Calculate V_r and L_r from above.
In stripping section:

Solve directly for L_s

$$L_s = L_r + qF$$

Solve for V_s:

$$\frac{V_r - V_s}{F} = 1 - q \tag{19.422}$$

Calculate L_s/B

9. Evaluate pinch compositions at the assumed temperatures of Step 7. If this temperature does not give a balance, other temperatures should be assumed and a balance sought as indicated below. Either of the following balances can be used, for convenience:

Rectifying:

$$\sum x_{Dri} = \sum x_n = \sum_{i=h,l,L'}\left(\frac{Dx_{Di}/V_r}{\alpha_i K_h - L_r/V_r}\right) = 1$$

$$\text{or } \sum x_n = \sum_{i=h,l,L'}\left(\frac{Dx_{Di}/V_r}{K_i - L_r/V_r}\right) = 1 \tag{19.423}$$

$$\text{or } \sum x_n = \sum_{i=h,l,L'}\left(\frac{x_{Di}}{(\alpha_i - 1)(L_r/D)_{min} + \alpha_i x_{hDI}/x_{hn}}\right) = 1$$

When the heavy key in the overhead is very small, less than 0.1 mol fraction, the last term of the denominator can be omitted.

$$\text{or} \sum x_n = \sum_{i=h,l,L'} \left(\frac{x_{Di}}{K_i + (K_i - 1)(L_r/V_r/D)_{min}} \right) = 1$$

Note that the calculations are only made for the heavy key, h; light key, 1; and all components lighter than it, L′. If there are split keys, the calculation is to include all components lighter than the heavy key.

Stripping pinch compositions:

$$\sum x_{Dsi} = \sum x_m = \sum_{i=H,h,l} \frac{Bx_{Bi}/V_s}{\frac{L_s}{V_s} - \alpha_i K_h} = 1$$

$$\text{or} \sum x_m = \sum_{i=H,h,l} \frac{Bx_{Bi}/V_s}{\frac{L_s}{V_s} - K_i} = 1 \tag{19.424}$$

$$\text{or} \sum x_m = \sum_{i=H,h,l} \frac{\alpha_i x_{Bi}}{(\alpha_l - \alpha_i)(L_s/B)_{min} + \alpha_i x_{lB}/x_{ld}} = 1$$

Because the second term of the denominator is usually negligible when the light key in the bottoms is very small (less than 0.1 mol fraction), this term is often omitted.

$$\text{or} \sum x_m = \sum_{i=H,h,l} \frac{x_{Bi}}{K_i + (1 - K_i) + (L_s/B)_{min}} = 1$$

Note that these calculations are made for the light key, 1; heavy key, h; and all components heavier than the heavy key, h. For split key systems, the calculations are made for all components heavier than the light key.

10. Calculate mol fraction ratio:

   a. Stripping pinch.

$$r_{ps} = \frac{\text{light key}}{\text{heavy key}}$$

   b. Rectifying pinch

$$r_{pr} = \frac{\text{light key}}{\text{heavy key}}$$

   c.

$$p = r_{ps}/r_{pr}$$

11. Calculate for each component in pinch.
Rectifying: apply only to components lighter than light key, i = L′

$$\frac{(\alpha_i - 1)\alpha_i}{\alpha_i}$$

Read from Figure 19.80 value of C_{ni} for each component.
Calculate for each component:

$$(C_{ni})\left(x_{i_{pr}}\right)$$

Sum these values:

$$\sum_{i=L'} C_{ni} x_{ip}$$

Stripping: apply only to components heavier than heavy keys, i = h.

$$(\alpha_1 - 1)\ (\alpha_1)$$

From Figure 19.80 value of C_{mi} for each component.
Calculate for each component:

$$C_{mi} \alpha_i x_{i_{ps}}$$

Sum these values:

$$\sum_{i=H} C_{mi}\, \alpha_i\, x_{i_{ps}}$$

12. Calculate

$$p' = \frac{1}{\left[1 - \sum C_n x_{ipr}\right]\left[1 - \sum C_m \alpha_i x_{ips}\right]} \tag{19.425}$$

If the two values of p are not very nearly equal, this requires a retrial with a new $(L/D)_{min}$, and a follow-through of the steps above.

When $r_{ps}/r_{pr} > p'$ the assumed $(L/D)_{min}$ is too high. Note that r_{ps}/r_{pr} changes rapidly with small changes in $(L/P)_{min}$; p′ changes slightly. When p = p′ the proper $(L/D)_{min}$ has been found. Colburn reports the method accurate to 1%. It is convenient to graph the assumed $(L/D)_{min}$ versus p and p′ in order to facilitate the selection of the correct $(L/D)_{min}$.

Example 19.37: Application of the Colburn Equation to Calculate the Minimum Reflux Ratio for a Multicomponent Mixture

A mixture of four components is as listed below, using n-butane as the base component.

Component	**Relative Vol.**	**x_f**	**$(S_r)_i = Dx_D/Bx_B$**
1	0.25	0.10	–
Heavy Key	0.50	0.225	0.05
n-butane	1.0	0.450	20.00
4	2.0	0.225	–

$$\mathrm{Dx}_{\mathrm{Di}} = \frac{(\mathrm{S_r})_i \mathrm{Fx}_{\mathrm{Fi}}}{1+(\mathrm{Sr})_i}$$

$$\mathrm{S_r} = \text{separation ratio} = \frac{\mathrm{Dx}_{\mathrm{Di}}}{\mathrm{Bx}_{\mathrm{Bi}}}$$

If all at top, $\mathrm{S_r} = 1$

If all at bottom, $\mathrm{S_r} = 0$

For component No. 3; Basis; 1 mol feed,

$$\mathrm{Dx}_{\mathrm{D3}} = \frac{(20)\left[(1)(0.450)\right]}{1+20} = 0.428$$

For component No. 2,

$$\mathrm{Dx}_{\mathrm{D2}} = \frac{(0.05)(1+0.225)}{1+0.05} = 0.01072$$

Because this is at minimum reflux, and adjacent keys system,

$$\mathrm{Fx}_{\mathrm{FL}} = \mathrm{Dx}_{\mathrm{DL}}$$

$$\mathrm{FX}_{\mathrm{FH}} = \mathrm{Bx}_{\mathrm{BH}}$$

Therefore, for component No.4, lighter than light key,

$$\mathrm{Fx}_{\mathrm{F4}} = (1)\ (0.225) = 0.225$$

Then, $\mathrm{Dx}_{\mathrm{D4}} = 0.225$

For component No. 1, heavier than heavy key, this component will not appear in the overhead.

Bottoms:

$$\mathrm{Fx}_{\mathrm{Fi}} = \mathrm{Dx}_{\mathrm{Di}} + \mathrm{Bx}_{\mathrm{Bi}}$$

$$\frac{\mathrm{Fx}_{\mathrm{Fi}}}{\mathrm{Bx}_{\mathrm{Bi}}} = \frac{\mathrm{Dx}_{\mathrm{Di}}}{\mathrm{Bx}_{\mathrm{Bi}}} + 1$$

$$\frac{\mathrm{Fx}_{\mathrm{Fi}}}{\mathrm{Bx}_{\mathrm{Bi}}} = (\mathrm{S_r})_i + 1$$

$$(\mathrm{S_r})_i = \frac{\mathrm{Dx}_{\mathrm{Di}}}{\mathrm{Bx}_{\mathrm{Bi}}}, \text{ by definition}$$

Then, $\mathrm{Bx}_{\mathrm{Bi}} = \dfrac{\mathrm{Fx}_{\mathrm{F}}}{(\mathrm{S_r})_i + 1}$

Component No. 1:

$$\mathrm{Fx}_{\mathrm{Fi}} = \mathrm{Bx}_{\mathrm{Bi}} = (1)\ (0.1) = 0.10$$

Component No. 2:

$$(S_r)_i = \frac{Dx_{D2}}{Bx_{B2}} = 0.05$$

Substituting in equation previously established,

$$Bx_{B2} = \frac{(1)(0.225)}{(0.05)+1} = 0.214$$

or, because Dx_{D2} has been calculated,

$$Bx_{B2} = \frac{Dx_{D2}}{(S_r)_i} = \frac{0.01072}{0.05} = 0.214$$

Component No. 3:

$$Bx_{B3} = \frac{Dx_{D3}}{(S_r)_i} = \frac{0.428}{20} = 0.0214$$

Component No. 4: This component will not be in the bottoms because it is lighter than the light key

Overhead:

$$Dx_{D1} + Dx_{D2} + Dx_{D3} + Dx_{D4} = D$$

$$0 + 0.01072 + 0.428 + 0.225 = D$$

$$D = 0.66372 \text{ mol overhead product/mol feed}$$

Composition of Overhead:

Component	**Dx_{Di}**	**Mol%**
1	0	0
2	0.01072	1.6
3	0.428	64.6
4	0.225	33.9
Total	0.66372	100.1

Composition of Bottoms:

Component	Bx_{Bi}	Mol%
1	0.10	29.9
2	0.214	63.8
3	0.0214	6.3
4	0	33.9
Total	0.3354	100.0

To have some idea of what value to use in Colburn's "exact" method for minimum reflux, use Colburn's "approximate" method to establish the order-of-magnitude of the minimum reflux:

$$\left(\frac{L}{D}\right)_{min} = \frac{1}{\alpha - 1}\left(\frac{x_D}{x_n} - \alpha \frac{x_{hD}}{x_{hn}}\right)$$

where

x_D and x_n = top and pinch compositions of a given light component
x_{hD} and x_{hn} = top and pinch compositions of the heavy key component
α = relative volatility of the given component with reference to the heavy key

Estimating Pinch Composition:

$$x_n(\text{approx}) = \frac{r_f}{(1 + r_f)(1 + \Sigma\alpha x_z)}$$

where

r_f = ratio of liquid composition of light to heavy key component on feed plate
x_n = mol fraction of a component in the liquid in the rectifying column pinch
x_z = mol fraction of a component in the liquid part of the feed where the feed is part vapor
r_f = 0.450/0.225 = 2.0
$\Sigma\alpha x_z$ = (0.5) (0.10) = 0.05
$\alpha_{\text{light/heavy key}} = \frac{1}{0.5} = 2.0$

x_n(approx) = 2.0/(1 + 2)(1 + 0.05) = 2/3.15 = 0.635

In terms of heavy key:

$$\alpha_{1/h} = 2.0$$

$$\alpha_{1/l} = 0.25$$

$$\alpha_{1/h} = (\alpha_{1/h})(\alpha_{1/l}) = (2.0)(0.25) = 0.5$$

$$\text{approx.}\left(\frac{L}{D}\right)_{min} = \frac{1}{(2.0-1)}\left(\frac{0.645}{0.635}\right)$$

$$x_{D3} = 0.646$$

$$(L/D)_{min} = 1.017 \text{ approx}$$

Now: Use Colburn's more detailed method:

$$\text{Assume}\left(\frac{L}{D}\right)_{\min} = 1.017$$

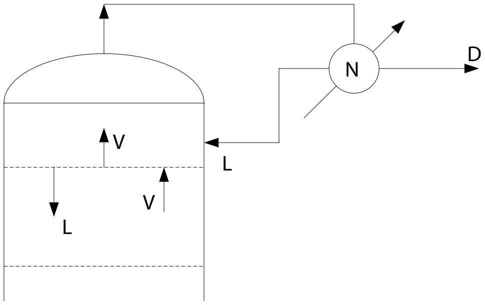
$$\left(\frac{L}{V}\right)_{\min} = \frac{1}{1+\left(\frac{D}{L}\right)_{\min}} = \frac{1}{1+\frac{1}{1.017}}$$

$$\left(\frac{L}{V}\right)_{\min} = \frac{1}{\frac{1.017+1}{1.017}} = 0.506$$

$$V = L + D$$

$$\frac{V}{L} = 1 + \frac{D}{L}$$

$$\frac{L}{V} = \frac{1}{1+\frac{D}{L}}$$



$$L_r = (0.506)\,(V_r)$$

and:

$$L_r = (1.017)\,(D_r)$$

Then, $(0.506)\,V_r = (1.017)\,(D_r) = (1.017)\,(0.66372)$

$$V_r = \frac{(1.017)(0.6637)}{0.506} = 1.332 \text{ mols/mol feed}$$

$$L_r = (1.017)\,(0.6637) = 0.674 \text{ mol/mol feed.}$$

The feed is a boiling point liquid from statement of problem:

$$q = 1.0$$

$$L_s = L_r + qF$$

Basis: 1 mol feed:

$$L_{s(min)} = 0.674 + (1)(1) = 1.674$$

$$\frac{L_r - V_s}{F} = 1 - q$$

$$\frac{1.332 - V_s}{1} = 1 - 1 = 0$$

$$V_s = 1.332$$

$$\frac{L_s}{V_s} = \frac{1.674}{1.332} = 1.255$$

Determine temperature of rectifying section pinch.

$$1 = \sum_{i=h,l,L} \frac{Dx_{Di}/V_r}{K_i - \frac{L_r}{V_r}}$$

Component	**Dx_{Di}**	**$\frac{Dx_{Di}}{V_r}$**	**K_B @ 118°F**	**α_i**	**$\alpha_i K_B$**	**$\frac{L_r}{V_r}$**	**$K_i - \frac{L_r}{V_r}$**	**$\frac{Dx_{Di}/V_r}{K_i - (L_r/V_r)}$**
h	0.01072	0.00805	1.06	0.50	0.53	0.506	0.024	0.333
l	0.428	0.321		1.0	1.06		0.554	0.580
L	0.225	0.169		2.0	2.12		1.614	0.104
Note: 0.225/1.332 = 0.169								Σ = 1.017
$K_i = \alpha_i K_B$								
B = reference								

Assume temperature at rectifying pinch. If the components were known, then the overhead dew point and bottoms bubble point could be determined, and from this an approximation could have been made of the pinch temperature. Because these cannot be calculated, trial-and-error must be used to obtain the correct pinch temperature.

Because 1.017 ≈ 1.0, continue calculations composition of rectifying pinch:

$$x_{pri} = \frac{Dx_{Di}/V_r}{K_i - L_r/V_r} = \frac{Dx_{Di}/V_r}{\alpha_i K_B - L_r/V_r}$$

Component	$\frac{Dx_{Di}/V_r}{K_i - L_r/V_r}$	Revised Mol Fraction
2 = h	0.333	0.333/1.017 = 0.328
3 = l	0.580	0.572
4 = L	0.104	0.102
	1.017	1.002

Determine temperature of stripping section pinch:

$$1 = \sum_{i=H,h,l} \frac{Bx_{Bi}/V_s}{\frac{L_s}{V_s} - K_i} = \frac{Bx_{Bi}/V_s}{\frac{L_s}{V_s} - \alpha_i K_B}$$

Component	Bx_{Bi}	$\frac{Bx_{Bi}}{V_s}$	L_s/V_s	K_B @ 130°F	α_i	$\alpha_i K_B$	$L_s / V_s - \alpha_i K_B$	$\frac{Bx_{Bi} / V_s}{(L_s / V_s) - \alpha_i K_B}$
		0.10						
1 = H	0.10	1.332 = 0.0752	1.255	1.23	0.25	0.308	0.947	0.0794
2 = h	0.214	0.1605			0.50	0.615	0.640	0.252
3 = l	0.0214	0.0160			1.0	1.23	0.025	0.640
								0.9714

Exact K_B must be between 1.23 and 1.24. Because there is a difference involved in calculation, the result is very sensitive to small changes in K.

Composition of stripping pinch:

Component	See Last Col. Above	Revised Mole Fraction
1 = H	0.0794	0.0794/0.9714 = 0.0817
2 = h	0.252	0.2585
3 = 1	0.64	0.659
	0.9714	0.9992

Calculate C_r for each component lighter than light key.

There is only one component lighter than light key in this example, #4

$$\left[\left(\frac{\alpha_l}{\alpha_h} - 1\right)\left(\frac{\alpha_l}{\alpha_L}\right)\right] = \left[\frac{1}{0.50} - 1\right]\left(\frac{1}{2}\right) = (2-1)\left(\frac{1}{2}\right) = 0.50$$

Reading curve, $C_r = 1.0 = C_n$ (see Figure 19.80)

Evaluate C_s for No. 1:

$$\left[\left(\frac{\alpha_l}{\alpha_h}-1\right)\alpha_H\right]=\left[\frac{1}{0.50}-1\right](0.25)=(2-1)(0.25)=0.25$$

Reading curve, $C_s = 0.875$ (see Figure 19.80).

Now, substitute into Colburn correlation, for check,

$$\frac{x_{prl}x_{psh}}{x_{prh}x_{psl}}=\left(1-\sum_L C_r x_{prl}\right)\left(1-\sum\frac{\alpha_{psH}}{\alpha_{psh}}C_S x_{psH}\right)$$

$$\frac{(0.572)(0.2585)}{(0.328)(0.659)}=(1-(1)(0.102))\left(1-\left(\frac{0.25}{0.50}\right)(0.875)(0.0817)\right)$$

$$0.684 = (1-0.102)\ (1-0.0318) = (0.898)\ (0.9682)$$

$$0.684 \neq 0.868.$$

Because left side of equation is smaller than right side, $\left(\frac{L}{D}\right)_{min}$ assumed was too large. Try a smaller value around 0.95. Right side of equation is not so sensitive to change.

where

L = all components lighter than light key

$\sum_L$ = sum of all components lighter than light key; L does not include light key

$\sum_H$ = sum of all components heavier than heavy key; H does not include heavy key

p = pinch

r = rectifying

s = stripping

l = light key

h = heavy key

H = all components heavier than heavy key, not including the heavy key

L = all components lighter than light key, not including the light key

C_r, C_s = empirical constants

19.38 Scheibel–Montross Empirical: Adjacent Key Systems: Constant or Variable Volatility [162]

This method has not found as much wide acceptance in use or discussion in the literature; nevertheless, it allows a direct approximate solution of the average multicomponent system with accuracy of 1–8% average. If the key components are less than 10% of the feed, the accuracy is probably considerably less than indicated. If a split key system is considered, Scheibel reports fair accuracy when the split components going overhead are estimated and combined with the light key, the balance considered with the heavy key in the L/D relation.

$$(L/D)_{\min} = \frac{1}{x_{lF} + \sum x_{FL}} \left[x_{lF}R' + \left(x_{hF} + \sum \frac{x_{FH}}{\dfrac{\alpha_l}{\alpha_H} - 1} + \sum \frac{x_{FL}}{\alpha_L}\left(1 + \frac{\alpha_l}{\alpha_L}\right) \right) \right] \tag{19.426}$$

where

x_{lF} = mol fraction of light key in feed
Σx_{FL} = sum of all mol fractions lighter than light key in feed
R' = pseudo minimum reflux
Σx_{FH} = sum of all mol fractions heavier than heavy key in feed
x_{hF} = mol fraction heavy key in feed
α_l = relative volatility of light key to heavy key at feed tray temperature
α_H = relative volatility of components heavier than heavy key at feed tray temperature
α_L = relative volatility of components lighter than light key at feed tray temperatures
x_{it} = mol fraction liquid at intersection of operating lines at minimum reflux (calculated or from graph)
x_{io} = mol fraction light key in overhead expressed as fraction of total keys in overhead

Pseudo minimum reflux:

$$R' = \frac{x_{io}}{(\alpha_i - 1)x_i} - \frac{(1 - x_i)(\alpha_i - 1)}{(1 - x_{io})\alpha_i} \tag{19.427}$$

When the overhead contains only a very small amount of heavy key, the second term in the equation may be neglected.

Intersection of operating lines at equilibrium curve:

$$x_{it} = \frac{\left[(\alpha_l - 1)(1 + m)\left(\dfrac{x_{lF}}{x_{lF} + x_{hF}}\right) - \alpha_l - m \pm \sqrt{\left\{ \left[(\alpha_l - 1)(1 + m)\left(\dfrac{x_{lF}}{x_{lF} + x_{hF}}\right) - \alpha_l - m \right]^2 + 4m(\alpha_l - 1)(1 + m)\left(\dfrac{x_{lF}}{xlF + x_{hF}}\right) \right\}} \right]}{2m(\alpha_l - 1)} \tag{19.428}$$

The proper value for x_{it} is positive and between zero and one. Actually, this is fairly straightforward and looks more difficult to handle than is actually the case.

Pseudo ratio of liquid to vapor in feed:

$$m = \frac{x_L - \sum xF_H}{x_v - \sum xF_L} = \frac{F_L - \sum F_H}{F_v - \sum F_L} \tag{19.429}$$

where

x_L = mole fraction of feed as liquid
x_v = mole fraction of feed as vapor
F_L = moles of liquid feed
F_v = moles of vapor feed
ΣF_H = total moles of components heavier than heavy key in feed
ΣF_L = total mole of components lighter than light key in feed.

Example 19.38: Scheibel–Montross Minimum Reflux [162]

A tower has the following all liquid feed composition:

Component	Feed Mol/hour	Overhead Mol/hour	Bottoms Mol/hour
A	30	30.	–
B (light key)	20	19.5	0.5
C (heavy key)	20	0.5	19.5
D	30	–	30.0
	100	50.0	50.0

Relative volatilities referenced to the heavy key, C:

$$\alpha_A = 4.0$$
$$\alpha_B = 2.0 = \alpha_1$$
$$\alpha_C = 1.0 = \alpha_h$$
$$\alpha_D = 0.5$$

$$\text{Calculate: } m = \frac{x_L - \sum xF_H}{x_v - \sum F_L} = \frac{1.0 - 0.30}{0 - 0.30} = -2.33$$

Intersection of Operating Lines:

$$x_{it} = \frac{\left[(2-1)(1-2.33)\left(\frac{0.2}{0.2+0.2}\right) - 2 - (-2.33) \pm \sqrt{\left\{\left[(2-1)(1-2.33)\left(\frac{0.2}{0.2+0.2}\right) - 2 - (-2.33)\right]^2 + 4(-2.33)(2-1)\left(\frac{0.2}{0.2+0.2}\right)\right\}}\right]}{2(-2.33)(2-1)}$$

$$x_{it} = 0.610, \text{ or } -0.459 \text{ (not acceptable)}$$

Pseudo minimum reflux ratio:

$$x_{lo} = \frac{19.5}{19.5+0.5} = 0.975$$

$$R' = \frac{0.975}{(2-1)(0.610)} - \frac{(1.0-0.975)(2)}{(1-0.610)(2-1)}$$

$$R' = 1.472$$

Minimum reflux ratio:

$$(L/D)_{min} = \frac{1}{0.2+0.3} \times \left[0.2(1.472) + (0.2+0.3)\left(\frac{0.3}{\frac{2}{0.5}-1}\right) + \frac{0.3}{4}\left(1+\frac{2}{4}\right)\right]$$

$$(L/D)_{min} = 0.912$$

19.39 Minimum Number of Trays: Total Reflux–Constant Volatility

The minimum theoretical trays at total reflux can be determined by the Fenske relation as previously given

$$S_m = N_{min} + 1 = \frac{\log\left(\frac{x_{Dl}}{x_{Dh}}\right)\left(\frac{x_{Bh}}{x_{Bl}}\right)}{\log \alpha_{avg}} \tag{19.430}$$

Note that N_{min} is the number of trays in the column and does not include the reboiler. When α varies considerably through the column, the results will not be accurate using the α_{avg} as algebraic average, and the geometric mean is used in these cases.

$$\alpha_{avg} = [(\alpha_t)(\alpha_b)]^{1/2}$$

For extreme cases, it may be necessary to calculate down from the top and up from the bottom until each section shows a fairly uniform temperature gradient between trays. Then the Fenske relation can be used for the remaining trays, using the conditions at the trays calculated as the terminal conditions instead of the actual overhead and bottoms.

19.39.1 Theoretical Number of Trays at Operating Reflux of a Multicomponent Mixture

The method of Gilliland [90] (Figure 19.49a or 19.76) is also used for multicomponent mixtures to determine theoretical number of trays at a particular operating reflux ratio or at various ratios. The Brown and Martin [163] curve of Figure 19.81 is also used in approximate the same manner, and produces essentially the same results, but is based on internal vapor and liquid flows.

The values needed to use the graph include

$$(L/V)_r = \frac{1}{1+(D/L)} \tag{19.431}$$

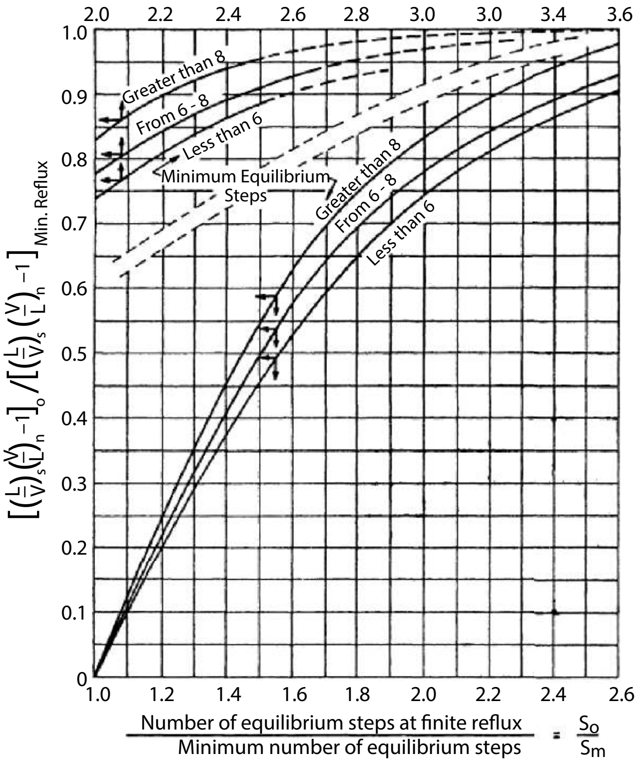


Figure 19.81 Brown and Martin: operating reflux and stages correlated with minimum reflux and stages. (Used and adapted by permission, Van Winkle, M., Oil and Gas Jour. V. 182, Mar. 23, (1953)).

where (D/L) = 1/ (L/D)

$$L_s = L_r + qF \tag{19.432}$$

$$L_r = (L/D)\,(D_r)$$

$$V_r = (L/D)\,D_r/(L/V) \tag{19.433}$$

$$V_s = V_r - F(1 - q) \tag{19.434}$$

Note than when $(L/D)_{min}$ is used as the starting basis, the L_r, L_s, V_r, V_s, and their ratios will be for the minimum condition, and correspondingly so when the operating reflux is used.

The combined Fenske–Underwood–Gilliland (FUG) method developed by Frank [84] is shown in Figure 19.79. This relates product purity, actual reflux ratio, and relative volatility (average) for the column to the number of equilibrium stages required. This does not consider tray efficiency. It is perhaps more convenient to use in the design of new columns than reworking existing columns and should be used only on adjacent-key systems.

Eduljee [164] evaluated published data and corrected relationships for determining the number of actual trays versus actual reflux with reasonably good agreement:

First attempt:

when

$$1.1 < R/R_m \leq 2.0$$

$$(S/S_m)(R/R_m) = 2.82$$

when

$$R/R_m > 2.0$$
$$(S/S_m)(R/R_m) = 0.7 + 1.06(R/R_m) \quad (19.435)$$

If the number of actual trays, S, calculates to be 27 or greater, then revert to the following for better accuracy:

Second attempt:

when

$$1.1 < R/R_m \leq 2.0$$
$$S = 2.71(R_m/R)(S_m) + 0.38 \quad (19.436)$$

when

$$R/R_m > 2.0$$
$$S = [0.67(R_m/R) + 1.02](S_m) + 0.38 \quad (19.437)$$

where:

n = number of theoretical trays in the rectifying section
R = reflux ratio (L/D)
S = number theoretical trays in the column, including reboiler

Subscript

m = minimum

The feed plate location, for either rectifying or stripping sections:

For R/R_m from 1.2 to 3.6:

$$(n/n_m)\,(R/R_m) = 1.1 + 0.9\,(R/R_m) \quad (19.438)$$

Hengstebeck [165] presents a technique for locating the feed tray by plotting.

Example 19.39: Operating Reflux Ratio

The minimum reflux ratio $(L/D)_{min}$ has been determined to be 1.017. Using the Brown and Martin graph [163], evaluate the theoretical number of trays at an operating reflux of 1.5 times the minimum. The minimum number of stages was determined to be 22.1 including the reboiler. See Figure 19.81.

The column will have a total condenser. Product rate D is 0.664 mol/mol feed, and the feed is a boiling point liquid.

Minimum values:

$$\left(\frac{L}{V}\right)_{min} = \frac{1}{1 + (D/L)_{min}} = \frac{1}{1 + 1/1.017} = 0.506$$

$$0.506Vr = 1.017\ (D_r) = 1.017\ (0.664)$$

$$V_r = 1.332 \text{ mol per mol of feed}$$

$$L_r = 1.017\ (0.664) = 0.674 \text{ mol/mol feed}$$

$$q = 1.0$$

$$L_s = (0.674) + (1)\ (1) = 1.674 \text{ mol/mol feed}$$

$$V_s = 1.332 - (1)\ (1 - 1) = 1.332 \text{ mol/mol feed}$$

$$\left(\frac{L}{V}\right)_s = \frac{1.674}{1.332} = 1.255$$

Operating values:

$$\text{Operating } (L/D)_o = (1.5)(1.017) = 1.525$$

$$\left(\frac{L}{V}\right)_o = \frac{1}{1 + 1/1.525} = 0.603$$

$$V_r = \frac{(1.525)(0.664)}{0.603} = 1.68 \text{ mol/mol feed}$$

$$L_r = (1.525)\ (0.664) = 1.013 \text{ mol/mol feed}$$

$$q = 1.0$$

$$L_s = 1.013 + (1)\ (1) = 2.013$$

$$V_s = 1.68 - (1)\ (1 - 1) = 1.68$$

$$(L/V)_s = 2.013/1.68 = 1.198$$

For graph:

$$\left[\left(\frac{L}{V}\right)_s \left(\frac{V}{L}\right)_r - 1\right]_o = 1.198\left(\frac{1}{0.603}\right) - 1 = 0.985$$

$$\left[\left(\frac{L}{V}\right)_s \left(\frac{V}{L}\right)_r - 1\right]_{min} = (1.255)\left(\frac{1}{0.506}\right) - 1 = 1.48$$

Read curve for "greater than 8" minimum equilibrium steps:

at 0.985/1.48 = 0.666
Curve reads (Figure 19.81): $S_o/S_M = 1.64$
Theoretical stages at reflux (L/D) = 1.525
$S_o = S_M (1.64) = 22.1 (1.64)$
$S_o = 36.2$ stages
Theoretical trays at the operating reflux (L/D) = 1.525
No = 36.2 − 1 (for reboi1er) = 35.2 trays in column

19.39.2 Actual Number of Trays

From the theoretical trays at operating reflux the actual trays for installation are determined:

$$N_{act} = N_o/E_o \tag{19.439}$$

The reboiler is considered 100% efficient, and likewise any partial condenser, if used. Therefore, the value N_o represents the theoretical trays or stages in the column proper, excluding the reboiler and partial condenser. E_o represents the overall tray efficiency for the system based upon actual test data of the same or similar systems, or from the plot of Figure 19.54, giving operating information preference (if reliable).

19.39.3 Estimation of Multicomponent Recoveries

Yaws *et al.* [166] present a useful technique for estimating overhead and bottoms recoveries with a very good comparison with tray-to-tray computer calculations. The procedure suggested uses an example from the reference with permission:

1. Plot relative volatility (α_i) and % desired recovery for LK and HK. Draw a straight line through these two points. The non-key component points will also be on this straight line.
2. Using α_i and the component distribution line, estimate % recovery of non-key components in distillate and bottoms.

From the references [139, 166]:

$$\log(d_i/b_i) = a + b\log\alpha_i \tag{19.440}$$

where

d_i = moles of component i in distillate
b_i = moles of component i in bottoms
α_i = relative volatility of component i
a, b = correlation constants
log (d_i/b_i) vs. log α_i gives a straight line (Figure 19.82).

By superimposing a Y_{iD} scale over the d_i/b_i scale,

$$d_i/b_i = Y_{iD}/Y_{iB}$$

where

Y_{iD} = % recovery of i in distillate
Y_{iB} = % recovery of i in bottoms
f_i = total moles of component i in distillate and bottoms

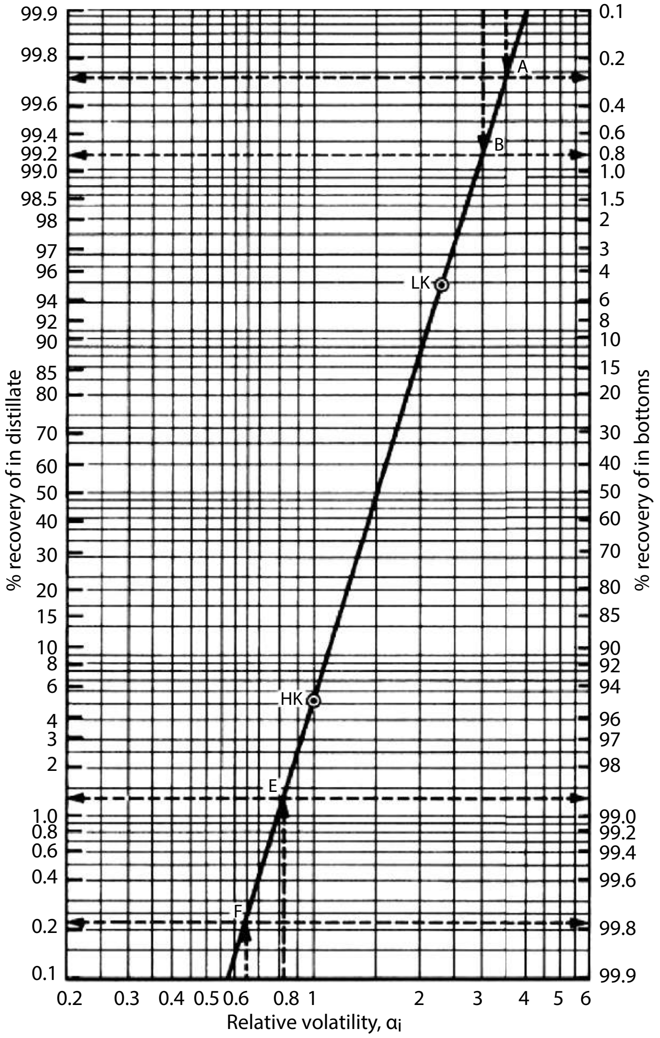


Figure 19.82 Estimation of recovery of non-key components using short-cut method of Yaws *et al.,* (Chem. Eng. Jan. 29, p. 101, (1979)).

$$\text{Then } Y_{iB} = 100 - Y_{iD} \tag{19.441}$$

$$\text{and } d_i/b_i = Y_{iD}(100 - Y_{iD}) \tag{19.442}$$

From Equation (19.441), Table 19.33 is constructed for selected values of d_i/b_i at various values of Y_{iD} from 99.9% to 0.1%.

19.39.4 Component Recovery Nomograph (Figure 19.82)

A nomograph is constructed by plotting d_i/b_i vs. α_i on log-log graph paper and then superimposing a Y_{iD} scale over the d_i/b_i scale, according to the values given in Table 19.33. The resulting nomograph, relating component recovered and component relative volatility, is given in Figure 19.82. This may be used to estimate component recovery in distillate and bottoms, as follows.

Example 19.40: Estimated Multicomponent Recoveries by Yaws' Method [166]

Component C is to be separated from Component D by distillation. A 95% recovery of both key components (LK, HK) is desired. Saturated-liquid feed composition and relative volatilities (at average column conditions) are given in Table 19.34.

Table 19.33 Table of Y_{iD} values for solving Yaws, Fang, and Patel short cut recoveries estimate.

Y_{iD}	d_i/b_i	Y_{iD}	d_i/b_i	Y_{iD}	d_i/b_i	Y_{iD}	d_i/b_i
99.9	999	96	24.0	40	0.6670	2	0.02040
99.8	499	94	15.7	30	0.4290	1.0	0.01010
99.6	249	92	11.5	20	0.2500	0.8	0.00806
99.4	166	90	9.00	15	0.1760	0.6	0.00604
99.2	124	85	5.67	10	0.1110	0.4	0.00402
99.0	99.0	80	4.00	8	0.0870	0.2	0.00200
98.5	65.7	70	2.33	6	0.0638	0.1	0.00100
98.0	49.0	60	1.50	4	0.0417		
97.0	32.3	50	1.00	3	0.0309		

Used by permission, Chem. Eng., Yaws, C. L., *et al.* Jan. 29 (1979), p. 101, All rights reserved.

Table 19.34 Material balance for estimated multicomponent distillation recoveries for Example 19.40 using method of Yaws, Fang, and Patel.

Feed component	**Composition, mol. Fr.**	**Relative volatility**	**Keys**
A	0.05	3.5	
B	0.20	3.0	
C	0.30	2.3	Light
D	0.25	1.0	Heavy
E	0.15	0.83	
F	0.05	0.65	
Distillate component	**Recovery desired, %**	**Recovery derived, %**	
A		99.72**	
B		99.20**	
C	95	-	
D	5	-	
E		1.30**	
F		0.22*	
Bottoms component	**Recovery desired, %**	**Recovery derived, %**	
A		0.28**	
B		0.80**	
C	5	-	

(*Continued*)

Table 19.34 Material balance for estimated multicomponent distillation recoveries for Example 19.40 using method of Yaws, Fang, and Patel. (***Continued***)

Feed component	Composition, mol. Fr.	Relative volatility	Keys
D	95	-	
E		98.70**	
F		99.78**	

*See calculations
**From Figure 10-99
Used by permission, Yaws, C. L., *et al.*, Chem. Eng., Jan. 29 (1979), p. 101, All rights reserved.

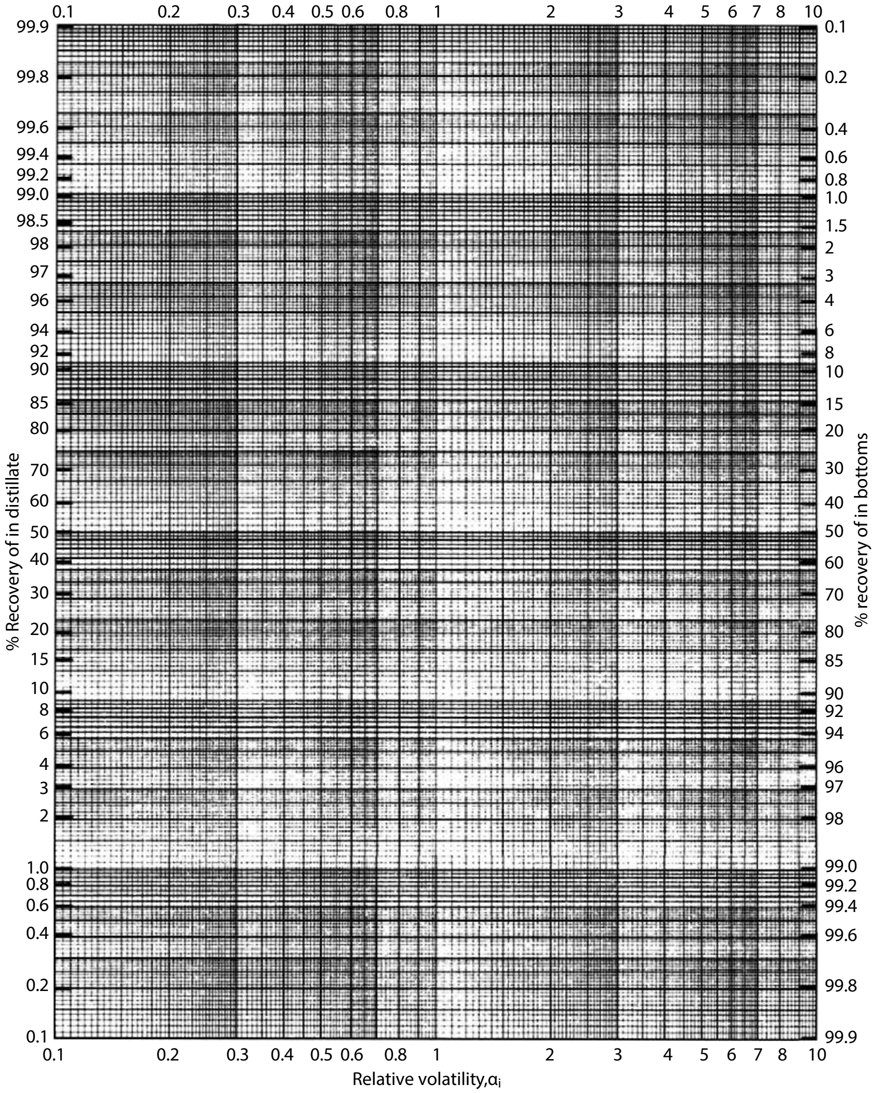


Figure 19.83 Working chart of Yaws *et al.*, short-cut method for multicomponent distillation for estimating component recovery in distillate and bottoms. (Used by permission, Yaws *et al.*, Chem. Eng. Jan. 29, p. 101, (1979)).

Using the graphical short-cut method for component distribution, estimate the recovery of non-key components in distillate and bottoms.

Solution:

1. α_i and % desired recovery are plotted for LK and HK ($\alpha_c = 2.3$, 95% recovery of C in distillate and $\alpha_D = 1$, 95% recovery of D in bottoms), as shown in Figure 19.82. See Figure 19.83 for working chart. A straight line is then drawn through the two points.
2. Using $\alpha_A = 3.5$, $\alpha_B = 3.0$, $\alpha_E = 0.83$, $\alpha_F = 0.65$, and the component distribution line, the recovery of non-key components is estimated. The results are shown in Table 19.33.

Table 19.35 [166] illustrates the good agreement between the proposed method with the tray-to-tray calculations for Case I-High Recovery: 95% LK recovery in distillate, 94% HK in bottoms; Case II, Intermediate Recovery: 90% LK recovery in distillate, 85% HK recovery in bottoms; and Case III Low Recovery: 85% LK recovery in distillate, 81 % HK recovery in bottoms. Figure 19.84 shows the Gilliland plot for multicomponent for Example 19.40.

Table 19.35 Comparison of Yaws, *et al.* short cut nomograph results vs. plate-to-plate calculations.

				Composition x_i	
				Distillate	**Bottoms**
	Component	**Nomograph**	**Plate to plate**	**Nomograph**	**Plate to plate**
Case I High recovery (16 trays)	A	0.0901	0.0901	0.0002	0.0002
	B	0.3588	0.3591	0.0026	0.0023
	C (LK)	0.5197	0.5190	0.0269	0.0278
	D (HK)	0.0271	0.0271	0.5271	0.5271
	E	0.0041	0.0045	0.3314	0.3308
	F	0.0002	0.0002	0.1118	0.1118
Case II Inter-mediate recovery (13 trays)	A	0.0879	0.0880	0.0016	0.0012
	B	0.3466	0.3464	0.0128	0.0120
	C (LK)	0.4814	0.4770	0.0683	0.0726
	D (HK)	0.0668	0.0682	0.4839	0.4835
	E	0.0155	0.0187	0.3218	0.3187
	F	0.0018	0.0018	0.1116	0.1120
Case III Low recovery (9 trays)	A	0.0866	0.0872	0.0034	0.0028
	B	0.3376	0.3395	0.0250	0.0227
	C (LK)	0.4561	0.4552	0.1015	0.1027
	D (HK)	0.0844	0.0839	0.4606	0.4610
	E	0.0308	0.0295	0.3016	0.3031
	F	0.0045	0.0046	0.1079	0.1077

Used by permission, Chem. Eng., Yaws, C. L., *et al.* Jan. 29 (1979), p. 101, all rights reserved.

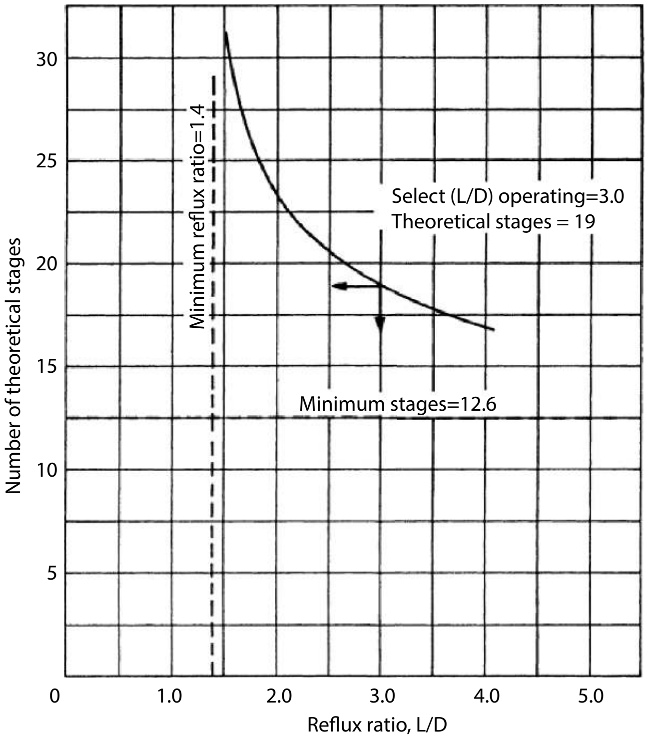


Figure 19.84 Gilliland plot for multicomponent for Example 19.40.

Example 19.41: Estimated Multicomponent Recoveries by Yaws' Method [166]

Table 19.36 [166] shows case 1 (High recovery of 11 components), 94.70% LK recovery and 95.52 % HK recovery and case 2 (low recovery of 13 components), 73.51% LK recovery and 63.43% HK recovery. Using the short-cut method for component distribution, estimate the recovery of non-key components in distillate and bottoms.

Solution

The Excel spreadsheet program is used to determine the recovery of non-key components in distillate and bottoms, using the thermal feed condition of saturated liquid, q = 1. The program (as illustrated earlier) determines the Underwood constant, the minimum reflux ratio minimum number of stages, actual reflux ratio, the number of theoretical plates, and the position of the feed plate. Tables 19.37 and 19.38 show the results of the Excel spreadsheet calculations. Comparison of these results with plate-to-plate calculations is shown in Table 19.39. There is good agreement between them in both cases. Further, Microsoft Excel spreadsheet Example 19.41a.xlsx is used to determine the number of theoretical stages using the Erbar-Maddox correlation and the Rusche Equation (19.387). The only drawback with the Excel spreadsheet program is that users will initially need to provide a default value of the Underwood constant (θ) and then use the Goal Seek from the Data-What-If Analysis menu to perform the iteration until the q-value (i.e., pseudo ratio of liquid to vapor for the feed) in the program is equal to zero. Furthermore, the components must be arranged so that the LK and HK are adjacent in the Excel spreadsheet program. The results are in good agreement with those from other simulations, and the Fortran developed program.

19.39.5 Shortcut Methods: Reflux Ratio and Stages

The shortcut methods allow determination of the number of theoretical plates as a function of reflux ratio, minimum number of plates, and minimum reflux. They are used to study the effect of reflux ratio on investment and operating costs with minimum computational effort. These methods are also useful for comparing other variables related to reflux ratio and number of stages. The Underwood [144] and Colburn [161] minimum reflux methods are used for more accurate calculation of minimum reflux, while the Brown–Martin [163] method is a suitable approximation. The Underwood

Table 19.36 Case 1.11 components 94.70 % LK recovery and 95.52 % HK recovery.

Component	f_i	α_i
M	0.06	2.5
N	0.03	2.3
O	0.05	2.1
P	0.09	1.75
Q	0.13	1.60
R (LK)	0.15	1.45
S (HK)	0.14	1.00
T	0.11	0.90
U	0.10	0.83
V	0.08	0.75
W	0.06	0.70

Table 19.36 Case 2. Low recovery (13 components) 73.51 % LK recovery and 63.43 % HK recovery.

Component	f_i	α_i
K	0.01	1.6
L	0.05	1.4
M	0.02	1.3
N	0.12	1.2
O	0.05	1.14
P	0.09	1.12
Q (LK)	0.16	1.10
R (HK)	0.17	1.00
S	0.04	0.98
T	0.11	0.94
U	0.09	0.90
V	0.06	0.80
W	0.03	0.65

Table 19.37 Excel spreadsheet calculations of Case I, 11 components 94.70% LK recovery and 95.52% HK recovery of Multicomponent System Fractionation.

Multicomponent System Fractionation										
11 components 94.70% LK recovery and 95.52% HK recovery										
Component Number	**Feed Moles**	**REL. VOL.α**	**Mol. Fraction xF**	**Product in Distillate, moles**	**Distillate %**	**Moles**	**xD**	**Bottoms %**	**Moles**	**xB**
1	0.06	2.5	0.06	0.999990791	99.9990791	0.05999945	0.11815379	9.209E-06	5.526E-07	1E-06
2	0.03	2.3	0.03	0.999965054	99.9965054	0.02999895	0.05907537	3.495E-05	1.048E-06	2E-06
3	0.05	2.1	0.05	0.999850296	99.9850296	0.04999251	0.09844766	0.0001497	7.485E-06	2E-05
4	0.09	1.75	0.09	0.99724253	99.724253	0.08975183	0.1767436	0.0027575	0.0002482	0.0005
5	0.13	1.6	0.13	0.988540831	98.8540831	0.12851031	0.25306866	0.0114592	0.0014897	0.003
6(LK)	0.15	1.45	0.15	0.947	94.7	0.14205	0.27973167	0.053	0.00795	0.0162
7(HK)	0.14	1	0.14	0.0448	4.48	0.006272	0.01235112	0.9552	0.133728	0.2717
8	0.11	0.9	0.11	0.008621567	0.86215669	0.00094837	0.00186758	0.9913784	0.1090516	0.2216
9	0.1	0.83	0.1	0.002376294	0.23762943	0.00023763	0.00046795	0.9976237	0.0997624	0.2027
10	0.08	0.75	0.08	0.000470687	0.04706866	3.7655E-05	7.4152E-05	0.9995293	0.0799623	0.1625
11	0.06	0.7	0.06	0.000156186	0.0156186	9.3712E-06	1.8454E-05	0.9998438	0.0599906	0.1219
	1		1		0.50780808	0.50780808	1		0.4921919	1

The feed condition, Q	1.0
The heavy key component number	7
Percentage recovery of the light key component in the distillate (%)	94.7
Percentage recovery of the heavy key component in the bottoms (%)	95.52
Total moles in the distillate, moles	0.5078
Total moles in the bottoms, moles	0.4922
Total moles in the feed, moles	1.0
Underwood constant (θ)	1.1578
Factor for the reflux ratio	1.15
Minimum reflux ratio	2.2977
Actual reflux ratio	2.64
Minimum number of stages	16
Number of theoretical plates in the column	37
The position of the feed plate	19

Table 19.38 Excel spreadsheet calculations of Case II, 13 components 73.51% LK recovery and 64.43% HK recovery of Multicomponent System Fractionation.

Multicomponent System Fractionation

13 components 94.70% LK recovery and 95.52% HK recovery

Component Number	**Feed Moles**	**REL. VOL.α**	**Mol. Fraction xF**	**Product in Distillate, moles**	**Distillate %**	**Moles**	**xD**	**Bottoms %**	**Moles**	**xB**
1	0.01	1.6	0.01	0.9992526	99.92525996	0.009992526	0.01914424	0.0007474	7.474E-06	1.56E-05
2	0.05	1.4	0.05	0.993284856	99.32848565	0.049664243	0.095149534	0.006715144	0.00033576	0.000702
3	0.02	1.3	0.02	0.977573861	97.7573861	0.019551477	0.037457814	0.022426139	0.00044852	0.000938
4	0.12	1.2	0.12	0.920940598	92.09405976	0.110512872	0.211726741	0.079059402	0.00948713	0.019846
5	0.05	1.14	0.05	0.833346421	83.33464213	0.041667321	0.079828584	0.166653579	0.00833268	0.017431
6	0.09	1.12	0.09	0.788802492	78.88024916	0.070992224	0.136010874	0.211197508	0.01900778	0.039762
7(LK)	0.16	1.1	0.16	0.7351	73.51	0.117616	0.225335312	0.2649	0.042384	0.088662
8(HK)	0.17	1	0.17	0.3657	36.57	0.062169	0.119106848	0.6343	0.107831	0.225569
9	0.04	0.98	0.04	0.292393566	29.23935663	0.011695743	0.022407358	0.707606434	0.02830426	0.059209
10	0.11	0.94	0.11	0.172096955	17.20969548	0.018930665	0.036268427	0.827903045	0.09106933	0.190506
11	0.09	0.9	0.09	0.092140943	9.214094323	0.008292685	0.01588759	0.907859057	0.08170732	0.170921
12	0.06	0.8	0.06	0.014348764	1.434876385	0.000860926	0.00164941	0.985651236	0.05913907	0.123712
13	0.03	0.65	0.03	0.000474415	0.047441482	1.42324E-05	2.72673E-05	0.999525585	0.02998577	0.062726
	1					0.521959914	1		0.47804009	1

The feed condition, Q	1.0
The heavy key component number	8
Percentage recovery of the light key component in the distillate (%)	73.51
Percentage recovery of the heavy key component in the bottoms (%)	64.43
Total moles in the distillate, moles	0.5219
Total moles in the bottoms, moles	0.4780
Total moles in the feed, moles	1.0
Underwood constant (θ)	1.0438
Factor for the reflux ratio	1.15
Minimum reflux ratio	5.1163
Actual reflux ratio	5.88
Minimum number of stages	16
Number of theoretical plates in the column	36
The position of the feed plate	13

Table 19.39 Short-cut versus plate-to-plate calculations.

Case 1. High recovery (11 components) 94.70 % LK recovery and 95.52% HK recovery

			% recovery	**in dist. $100d_i/f_i$**	**% recovery**	**in btms. $100b_i/f_i$**	**Absolute deviation**
Component	**f_i**	**α_i**	**Short-cut**	**Plate-to-Plate**	**Short-cut**	**Plate-to-plate**	**|Δ|, %**
M	0.06	2.5	99.9991	99.999	0.0009	0.001	0.0001
N	0.03	2.3	99.9965	99.996	0.0035	0.004	0.0005
O	0.05	2.1	99.9850	99.986	0.0150	0.014	0.0010
P	0.09	1.75	99.7243	98.698	0.2757	0.302	0.0263
Q	0.13	1.6	98.8541	98.826	1.1459	1.174	0.0281
R (LK)	0.15	1.45	94.7	94.722	5.3000	5.278	0.0220
S (HK)	0.14	1.0	4.48	4.501	95.520	95.499	0.0210
T	0.11	0.9	0.8622	0.83	99.1378	99.170	0.0322
U	0.10	0.83	0.2376	0.221	99.7624	99.779	0.0166
V	0.08	0.75	0.0471	0.042	99.9529	99.958	0.0051
W	0.06	0.7	0.0156	0.013	99.9844	99.987	0.0026
							Avg. 0.0141 %

Case 2. (13 components) 73.51 % LK recovery and 64.43 % HK recovery

			% recovery	**in dist. $100d_i/f_i$**	**% recovery**	**in btms. $100b_i/f_i$**	**Absolute deviation**
Component	**f_i**	**α_i**	**Short-cut**	**Plate-to-Plate**	**Short-cut**	**Plate-to-plate**	**|Δ|, %**
K	0.01	1.6	99.9253	99.911	0.0747	0.089	0.0143
L	0.05	1.4	99.3285	99.274	0.6715	0.726	0.0545
M	0.02	1.3	97.7574	97.745	2.2426	2.255	0.0124
N	0.12	1.2	92.0941	92.264	7.9059	7.736	0.1699
O	0.05	1.14	83.3346	83.677	16.6654	16.323	0.3424
P	0.09	1.12	78.8802	79.228	21.1198	20.772	0.3478
Q (LK)	0.16	1.10	73.5100	73.896	26.4900	26.104	0.3860
R (HK)	0.17	1.00	36.5700	36.264	63.4300	63.736	0.3060
S	0.04	0.98	29.2394	28.827	70.7606	71.173	0.4124
T	0.11	0.94	17.2097	16.818	82.7903	83.182	0.3917
U	0.09	0.90	9.2141	9.031	90.7859	90.969	0.1831
V	0.06	0.80	1.4349	1.491	98.5651	98.509	0.0561

(*Continued*)

Case 2. (13 components) 73.51 % LK recovery and 64.43 % HK recovery. (*Continued*)

			% recovery	in dist. $100d_i/f_i$	% recovery	in btms. $100b_i/f_i$	Absolute deviation
Component	f_i	α_i	**Short-cut**	**Plate-to-Plate**	**Short-cut**	**Plate-to-plate**	**$\lvert\Delta\rvert$, %**
W	0.03	0.65	0.0474	0.064	99.9526	99.936	0.0166
							Avg. 0.207 %

Note: Case 1: N = 31, N_f = 17, D = 50.8, R = 3.05
Case 2: N = 28, N_f = 15, D = 52.2, R = 8.5
Source: Yaws, C. L., *et al.*, Chem. Eng., p 101, Jan. 299, 1979.

[88] or Fenske methods are useful for estimating minimum number of stages, and the Brown–Martin [163] and Gilliland [90] correlations relate minimum reflux, minimum number of plates, operating reflux, and number of theoretical plates.

The minimum number of stages at total reflux as given by Fenske's equation is

$$N_m = \frac{\log\left[\left(\frac{x_{LK}}{x_{HK}}\right)_D \left(\frac{x_{HK}}{x_{LK}}\right)_B\right]}{\log\left(\frac{\alpha_{LK}}{\alpha_{HK}}\right)} \tag{19.183}$$

The Fenske's equation has been shown to be rigorous if

$$(\alpha_{LK}/\alpha_{HK})_{avg} = \sqrt[N]{\alpha_{LK}/\alpha_{HK,1}\ \alpha_{LK}/\alpha_{HK,2}\ldots\ldots\ldots\ldots\ldots\ldots\ldots\alpha_{LK}/\alpha_{HK,N}} \tag{19.443}$$

where $(\alpha_{LK}/\alpha_{HK})_{avg}$ is usually obtained from one of the following approximations:

1. Evaluate $(\alpha_{LK}/\alpha_{HK})_{avg}$ at $T_{avg} = (T_{top} + T_{bot})/2$ (19.443a)
2. $(\alpha_{LK}/\alpha_{HK})_{avg} = (\alpha_{top} + \alpha_{bot})/2$ (19.443b)
3. $(\alpha_{LK}/\alpha_{HK})_{avg} = \alpha$ at feed tray temperature (19.443c)
4. $(\alpha_{LK}/\alpha_{HK})_{avg} = \sqrt{\alpha_{top}\,\alpha_{bot}}$ (19.443d)
5. $(\alpha_{LK}/\alpha_{HK})_{avg} = \sqrt[3]{\alpha_{top}\,\alpha_{mid}\,\alpha_{bot}}$ (19.443e)

Listed below is the preferred $(\alpha_{LK}/\alpha_{HK})_{avg}$ by investigators, which is based upon the above conditions in Equation (19.443a–e).

Method	**Recommendation**
1	Maddox [130, 59]
2	van Winkle [82]
3 or 4	King [167]
4	Wankat [99], Ludwig [81]

Douglas [168] proposed a criterion for testing the relative volatility approximation as

$$\frac{\alpha_{top}-\alpha_{bot}}{\alpha_{top}+\alpha_{bot}} \le 0.1\ln\left(\frac{\alpha_{top}+\alpha_{bot}}{2}\right) \tag{19.444}$$

When the above inequality is obeyed, the relative volatility is reasonably constant throughout the column, and the simpler approximations such as 2 or 4 are suitable. The Fenske's equation applies to any pair of components and not necessarily to the light key and heavy key.

19.40 Smith–Brinkley (SB) Method [169]

Smith and Brinkley [169] developed a method for estimating the distribution of components in multicomponent separation processes. The method is based on an analytical solution of the finite-difference equations that can be written for multistage separation processes when stages and interstage flow rates are known. The Fenske–Underwood–Gilliland (FUG) method combines Fenske's total reflux equation and Underwood's minimum reflux equation with a graphical correlation by Gilliland. The latter relates actual column performance to total and minimum reflux conditions for a specified separation between two key components. The method can be used for extraction and absorption processes as well as distillation. The SB and FUG methods are rating and design methods, respectively, and both work best when mixtures are nearly ideal. Only the equations for distillation are presented here, as the derivation of the equations are given by Smith and Brinkley [169] and Smith [49]. For any component i (suffix i omitted in the equation for clarity)

$$\frac{b}{f}=\frac{\left(1-S_r^{N_r-N_s}\right)+R(1-S_r)}{\left(1-S_r^{N_r-N_s}\right)+R(1-S_r)+GS_r^{N_r-N_s}\left(1-S_r^{N_r+1}\right)} \tag{19.445}$$

where

b/f = the fractional split of the component between the feed and the bottoms
N_r = number of equilibrium stages above the feed.
N_s = number of equilibrium stages below the feed
S_r = stripping factor, rectifying section = $K_i V/L$
S_s = stripping factor, stripping section = $K_i' V'/L'$
V and L = total molar vapor and liquid flow rates, and the superscript ′ denotes the Stripping section
G = Depends on the condition of the feed.

If the feed is mainly liquid:

$$G_i=\frac{K_i'}{K_i}\frac{L}{L'}\left[\frac{1-S_r}{1-S_s}\right]_i \tag{19.446}$$

and the feed stage is added to the stripping section.

If the feed is mainly vapor:

$$G_i=\frac{L}{L'}\left[\frac{1-S_r}{1-S_s}\right]_i \tag{19.447}$$

K_i and K'_i are determined at the effective top and bottom section temperatures. If the temperature profile is available (e.g., from a computer simulation), the effective temperature is the arithmetic average of all tray temperatures in the column section. Alternatively, an arithmetic average of the feed-stage and end-stage temperatures is also used.

$$t_r = \frac{t_{top} + t_{above\ feed}}{2} \quad (19.448)$$

$$t_s = \frac{t_{below\ feed} + t_{bot}}{2} \quad (19.449)$$

Application

The Smith–Brinkley method rates a column using the reverse sequence of steps. It takes the number of stages, reflux ratio, and actual feed location and determines the product compositions. Once a base case is given, small changes in variables can readily be made. For example, for a change in feed temperature, half the change is added to t_r and the other half to t_s. For a change in reflux rate, the extra liquid is added to S_s and S_r. If the changes are large, some trial-and-error calculation is required, in which t_s and t_r are varied. This is similar to an operator who adjusts the control temperature by trial and error until the desired end products are achieved. These features make the Smith–Brinkley method valuable for on-line optimization (e.g., using computer or microprocessor control).

Equation (19.446) is for a column with a total condenser. If a partial condenser is used, the number of stages in the rectifying section should be increased by one. The procedure for using the Smith–Brinkley method is as follows [169]:

1. Estimate the flow rates, L, V and L′, V′ from the specified component separations and reflux ratio.
2. Estimate the top and bottom temperatures by calculating the dew and bubble points for assumed top and bottoms compositions.
3. Estimate the feed point temperature.
4. Estimate the average component K values in the stripping and rectifying sections.
5. Calculate the values of $S_{r,i}$ for the rectifying section and $S_{s,i}$ for the stripping section.
6. Calculate the fractional split of each component and hence the top and bottom compositions.
7. Compare the calculated with the assumed values and check the overall column mass balance.
8. Repeat the calculation until a satisfactory material balance is obtained. The usual procedure is to adjust the feed temperature up and down till a satisfactory balance is obtained.

The SB is a rating method, which is suitable for determining the performance of an existing column rather than a design method, as the number of stages must be known. It can be used for design if estimation of the number of stages is first accomplished by some other method. Equation (19.445) is then used to determine the top and bottom compositions. The estimated stages can thus be adjusted and the computations repeated until the required specifications are achieved. The Geddes–Hengstebeck method (as illustrated earlier) is easier to use and gives satisfactory results for preliminary design in estimating the component splits. Examples of the application of the Smith–Brinkley method are given by Smith [49].

Commercial process simulators have a wide range of distillation models with varying degree of sophistication. It is essential that the model is chosen for the problem type, the degree of information available, and the level of detail required in the solution. The simplest distillation models are shortcut models, which use the Fenske–Underwood–Gilliland (FUG) or Winn–Underwood–Gilliland (WUG) method to determine the minimum reflux ratio and the number of stages. Alternatively, they determine the required reflux given a number of stages or the required number of stages for a given reflux ratio. The shortcut methods can also estimate the condenser and reboiler duties and determine the optimum feed stage. The minimum information required to specify a shortcut distillation model is

- The component recoveries of the light and heavy key components.
- The condenser and reboiler pressures.
- Whether the column has a total or partial condenser.

There are instances where the designers can specify the purities of the light and heavy key components in the distillate and bottoms, respectively. However, care must be taken when using purity as a specification, as it is easy to specify infeasible purities or combinations of purities.

The easiest way to use a shortcut distillation model is to start by estimating the minimum reflux ratio and number of stages. The optimum reflux ratio is usually between 1.05 and 1.25 times the minimum reflux ratio; however, $R = 1.15R_{min}$ is often used as initial estimate. Once the reflux ratio is specified, the number of stages and optimum feed stage can be determined. The shortcut model results can start as a basis for rigorous distillation simulation. Shortcut methods can be used to initialize fractionation columns with multiple products, as they are versatile and are solved rapidly. They do not give an accurate prediction of the distribution of non-key components, and they do not perform well when there is significant liquid-phase non-ideality. The main disadvantage of the shortcut models is that they assume constant relative volatility, which is usually calculated at the feed condition. Such models should not be used, if there is significant liquid or vapor phase nonideality, as this assumption does not hold at all well.

Minimum Reflux Ratio and Minimum Number of Stages by Simulation

Determining the minimum reflux ratio and the minimum number of stages, one develops a reflux stage plot and extrapolates from it. To develop this plot, simulation runs are performed at a different number of stages while keeping the material balance, product compositions and the ratio of the feed stage to the number of stages constant. The reflux ratio is varied; then, a plot of the number of stages versus reflux or reflux ratio is plotted. The curve is extrapolated asymptotically to an infinite number of stages to obtain the minimum reflux ratio and asymptotically to an infinite reflux ratio to obtain the minimum number of stages.

Optimization of the Feed Stage by Simulation

Simulation runs are performed at different feed positions to determine the optimum feed stage. Here, the material balance, reflux ratio and the total number of stages are kept constant. The two main plots can be created, namely: the McCabe-Thiele diagram, and the concentration versus feed stage diagram. The McCabe-Thiele diagram is plotted using the mole fraction data calculated for each stage by the simulation. The equilibrium data and the operating lines are determined from this data. The McCabe-Thiele diagram then shows how an optimum feed stage versus a non-optimum feed stage resembles when using the simulation data.

19.41 Retrofit Design of Distillation Columns

Since distillation is an energy-intensive process requiring considerable capital investment, retrofit of distillation columns is often carried out in preference to installation of new equipment. Retrofits aim to reuse the existing equipment more effectively in order to increase profit and to reduce cost, for example, by increasing a unit's capacity. When such a revamp is carried out, it is important to make effective utilization of the existing equipment. If the capacity of a system having two existing columns needs to be increased, then instead of replacing the existing columns with new ones, a third column might be added to the existing columns and the new system reconfigured. In carrying out a retrofit study, retrofit models are employed to fix the existing distillation design, as they are quicker to solve and more robust for optimization, especially when all design variables are being considered simultaneously. They can also be combined with detailed heat-integration models for improving the energy efficiency of distillation systems. Heat integration of distillation columns is reviewed later in the book.

Gadalla *et al.* [170] presented retrofit shortcut models for the design of reboiled and steam stripped distillation columns. These models are particularly valuable for evaluating retrofit design options and for improving the performance of existing distillation systems. The models are based upon a modified Underwood method, the Gilliland correlation, the Kirkbride equation, the Fenske equation, and the material balances. The models fix both column configurations and operating conditions such as the steam flow rates, and then calculate the product flow rates, temperatures, and compositions, and the various heat duties.

These models account for the changes in relative volatility and molal overflow through the column, overcoming the underlying limitations of the previous shortcut models. The models are reliable for very complex configurations, including a large number of components such as crude oil distillation. The models also apply to other distillation applications, such as naphtha fractionation and petrochemical separation. The retrofit models are suited for simple distillation columns, sequences of simple distillation columns, and complex distillation configurations including columns with side-strippers and side-rectifiers. They provide a basis for optimizing and improving operating conditions of existing columns for energy related, economic, and environmental benefits. They can be applied to estimate the additional heating and cooling requirements for increased throughput to an existing distillation process. Detailed presentation of these methods is given by Gadalla *et al.* [170].

Example 19.42: Shortcut Method Using Honeywell UniSim® Simulation Software

Estimate the number of ideal stages needed in the butane-pentane splitter defined by the compositions given in the table below. The column will operate at a pressure of 120 psia (830kPa), with a reflux ratio of 2.5. The feed is at its boiling point (q = 1).

Component	x_F	**Feed (f_i)**
Propane, C_3	0.05	5
Isobutane, iC_4	0.15	15
Normal butane, nC_4	0.25	25
Isopentane, iC_5	0.20	20
Normal pentane, nC_5	0.35	35

For a specification of not more than 1 mol of the light key in the bottoms product, and not more than 1 mol of the heavy key in the distillate product. Use the Peng–Robinson VLE package.

Solution

The Honeywell UniSim® Design R460.2 simulation (Example 19.42) package is used to separate n-butane and i-pentane from the components in the feed entering at its boiling point condition and at 120 psia. The shortcut model in UniSim requires less data than the rigorous model; only five input parameters are required to solve the problem. These are

1. Two key components: The light key (LK) in the bottoms and the heavy key (HK) in the distillate. The light key is defined as the most volatile component (e.g., n-butane) to appear at the bottom of the column, and the heavy key is defined as the least volatile component (e.g., i-pentane) to appear at the top of the column. The concentrations of the light and heavy key components must be specified.
   NB: Setting these values too low will result in a large (tall) column with a high reflux, while choosing high values may result in poor product quality and a weak separation.
2. Two pressure specifications: The reboiler and condenser pressures should be specified, as the estimates of condenser and reboiler pressures have a direct relationship on the temperatures that UniSim will estimate in these locations. The reboiler pressure should be greater than the condenser pressure.

NB: Until the user is satisfied with making estimates about the values, an initial guess of the pressures is required, and subsequent checks that the temperature estimates are reasonable need to be carried out. The reflux ratio R: the actual reflux ratio should be specified.

Additionally, a fully defined feed stream is required to be set up. Once all the required data are entered, UniSim® simulation software will calculate the remaining column parameters, which can be found on the **Performance tab**. The user should then check the temperature estimates on this page and, if necessary, adjust the reboiler and condenser pressures.

The following screen shots from UniSim® software illustrate the procedure of carrying out this simulation process.

Figures 19.85–19.103 provide snapshots of UniSim Design simulation of Example 19.42.

Step 1. Double click on UniSim icon to display the simulation window as shown in Figure 19.85.
Step 2. Open the file as a new case as in Figure 19.86 named simulation Basis Manager. Click on the **Add** button to select the components.
Step 3. Click on the Add Pure button to select the components as highlighted in Figure 19.87. Close the window.
Step 4. The Simulation Basis Manager window appears as shown in Figure 19.88. Click on **Fluid Pkgs tab** as shown in Figure 19.88. Then click on the **Add** button.
Step 5. Choose the required property package from the **Property Package Selection** menu. For this example, choose **Peng Robinson** as shown in Figure 19.89. Select **EOSs** from **Property Package Filter** and close the window.
Step 6. Close the window and return to the Simulation Basis Manager window as shown in Figure 19.90. Click on **Enter Simulation Environment** button as shown in Figure 19.90.
Step 7. Here, the Process Flow Diagram (PFD) window appears; then from the palette, double click on the **blue** arrow and place it on the PFD window.
Next, select the shortcut distillation icon in the palette by double clicking on it and place this on the PFD window as shown in Figure 19.91. This is named **T-100**. Double clicking on the arrow (→1) gives the worksheet window.
Step 8. Enter the **Stream Name, Pressure, Molar flow** from **Conditions** menu in the Worksheet of the Feed window as shown in Figure 19.92.
Step 9. Next, select **Composition menu** from the Worksheet in the Feed window. This provides the five components C_3–C_5 that were previously selected for this example.
Step 10. Select **Mole fraction** tab from **Input Composition for Stream**: Feed window and input the data values of the components as shown in Figures 19.93 and 19.94, respectively. Click on **OK** button to close the window.
Step 11. The **total Mole fraction is 1.0** as shown in Figure 19.95. Close this window and double click on the column icon on the PFD window.
Step 12. Double clicking on the **Red icon** of the column shows the Column window as shown in Figure 19.96. Click on the **Design tab** and enter the different streams: Feed, Distillate, Bottoms, Reb-Q, Cond-Q. Choose Liquid button for Top Product Phase from **Connections tab** as shown in Figure 19.96.
Step 13. Next, select the Parameter menu from Design window. Select light key in Bottoms and its Mole fraction tab, and Heavy key in Distillate and its corresponding Mole fraction tab as shown in Figure 19.97. Notice the highlighted red components displayed **"Unknown Key Components"** at the bottom of the window.
Step 14. Input the Parameter values and notice the highlighted red comments displayed **"Unknown Pressure"** as shown in Figure 19.98.
Step 15. Next, select **Parameters** from the **Design** window. Select Light Key in Component name from the list in Component **"Light Key in Bottoms"** (e.g., n-Butane) and enter its value under Mole fraction label. Select Heavy Key component name from the list in Component **"Heavy Key in**

Distillate" (e.g., i-Pentane) and enter its corresponding value under Mole fraction label. Enter Condenser and Reboiler Pressure (**NB: The value of Reboiler pressure is greater than the value of Condenser pressure**.). Enter the value of External Reflux Ratio. This gives a highlighted green comment **"OK"** as shown in Figure 19.99.

Step 16. The PFD window shows the results of shortcut simulation with **Feed, Distillate** and **Bottoms** and their corresponding values of **Temperature, Pressure** and **Molar flow rate** as shown in Figure 19.100.

Step 17. A summary of the printout of the data sheet can be obtained by **right clicking** on the column icon in Figure 19.101 to display a menu with **Print Datasheet** as shown in Figure 19.101. Then click on the **Print Datasheet** menu.

Step 18. Click on **Print Datasheet** to display "Select Data block(s) to Print for Shortcut Column" window as shown in Figure 19.102. Click on **Preview** button to view the Shortcut Column datasheet results or alternatively on **Print** button to print the hard copy of the results.

Step 19. Display results of the Shortcut Column simulation.

Table 19.40 shows the results of the Excel spreadsheet of Example 19.42 and the summary of the results of the shortcut method obtained from both the UniSim® Design R460.2 simulation software and the Excel spreadsheet program (Example 19.42.xlsx) is shown in Table 19.41. The results show that there is good agreement between them both.

Figure 19.103 shows the simulation results of Example 19.42.

19.42 Tray-by-Tray for Multicomponent Mixtures

Rigorous tray-by-tray computations for multicomponent mixtures of more than three components can be very tedious, even when made omitting a heat balance. Simulation software packages are adept at dealing with detail and several computation methods are in use.

The direct-solution method of Akers and Wade [171] is among several that attempts to reduce the number of trial-and-error solutions. This has been accomplished and has proven quite versatile in application. The adaptation outlined here modifies the symbols and rearranges some terms for convenient use by the designer [172]. Dew point and bubble point compositions and the plate temperatures can be determined directly. Constant molal overflow is assumed, and relative volatility is held constant over sections of the column.

Rectifying section: reference component is heavy key, x_h

$$\left(\frac{x_i}{x_h}\right)_n = \frac{1}{\alpha_i}\left[\frac{(L/D)(x_i)_{n+1} + x_{Di}}{(L/D)(x_h)_{n+1} + x_{Dh}}\right] \tag{19.450}$$

$$\Sigma x_i = 1.0 \text{ (including } x_h)$$

$$\sum\left(\frac{x_i}{x}\right)_n = \left(\frac{1}{x_h}\right)_n$$

The compositions of each component are obtained from $(x_i/x_h)_n$ ratio.

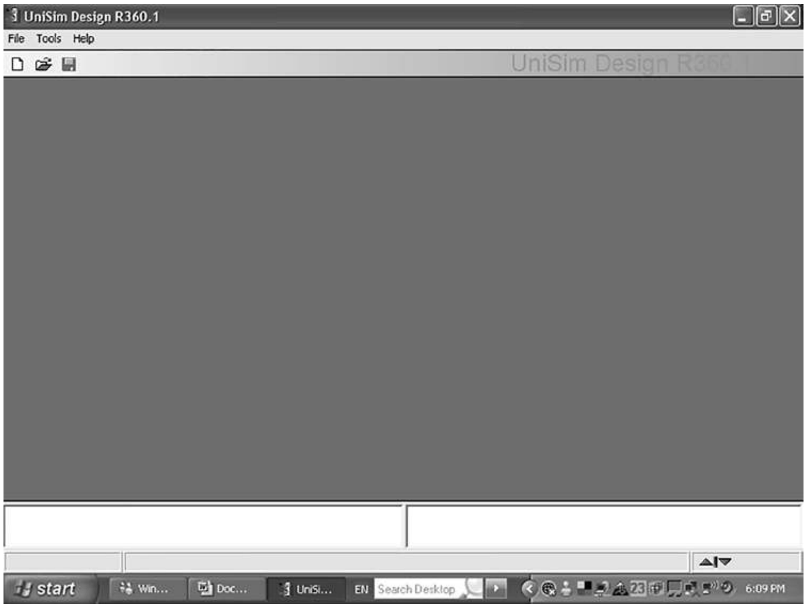


Figure 19.85 Snapshot of UniSim design Window. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).

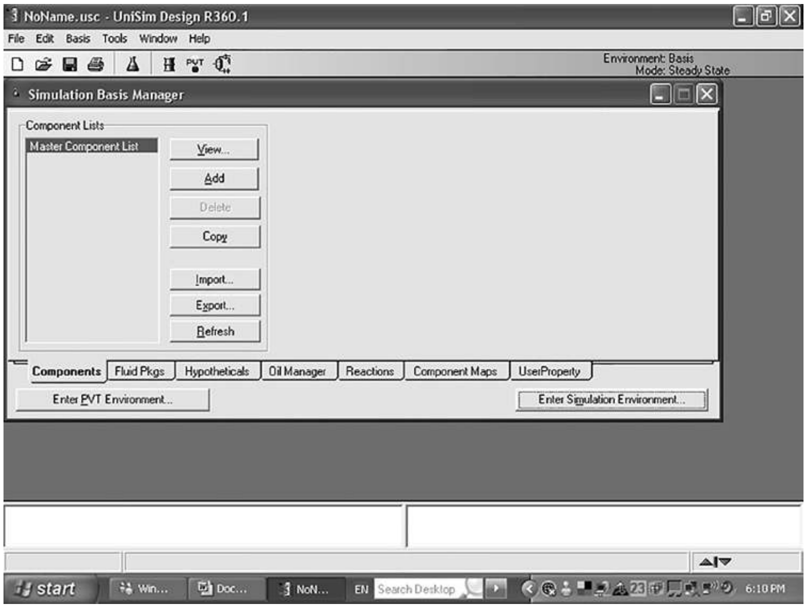
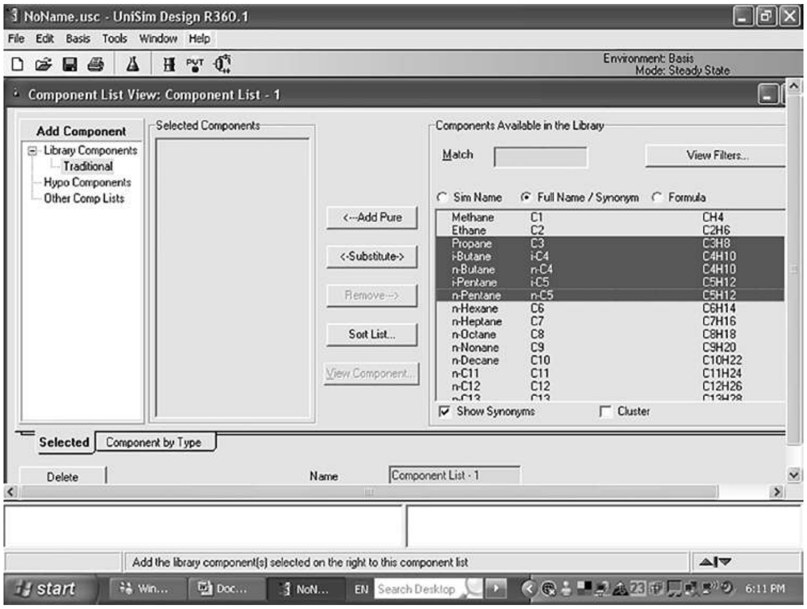


Figure 19.86 Snapshot window of Simulation Basis Manager. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).

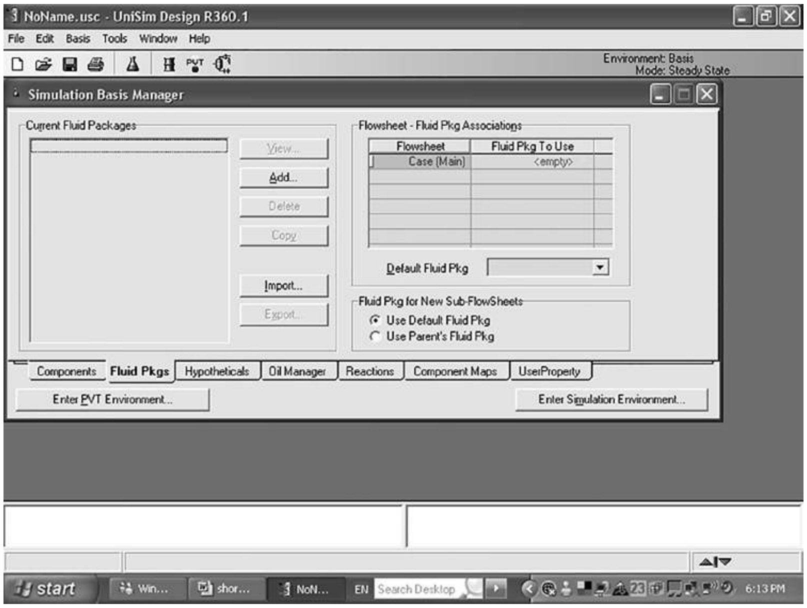


Methane	C1	CH4
Ethane	C2	C2H6
Propane	C3	C3H8
i-Butane	i-C4	C4H10
n-Butane	n-C4	C4H10
i-Pentane	i-C5	C5H12
n-Pentane	n-C5	C5H12
n-Hexane	C6	C6H14
n-Heptane	C7	C7H16
n-Octane	C8	C8H18
n-Nonane	C9	C9H20
n-Decane	C10	C10H22
n-C11	C11	C11H24
n-C12	C12	C12H26
n-C13	C13	C13H28

Name: Component List - 1

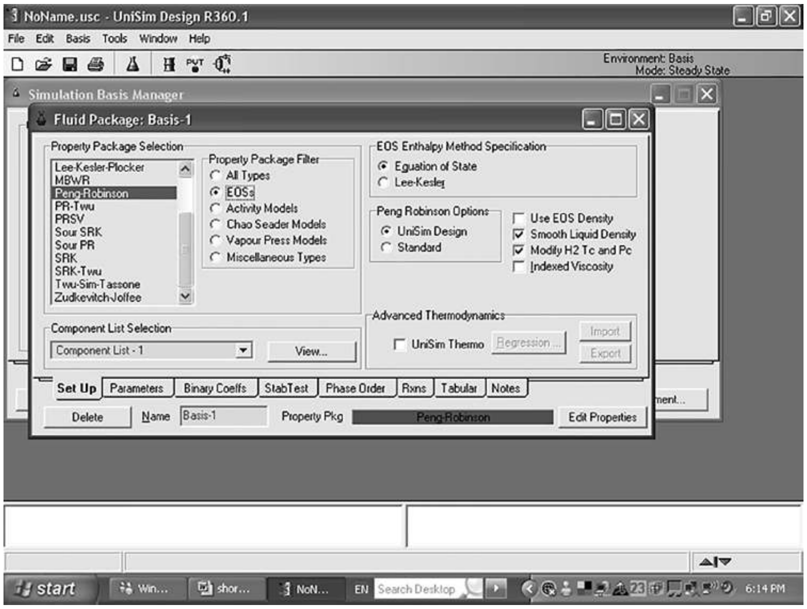
Add the library component(s) selected on the right to this component list

Figure 19.87 Snapshot of the component List View window showing the components. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).



Flowsheet	Fluid Pkg To Use
Case (Main)	<empty>

Figure 19.88 Snapshot of Simulation Basis Manager window. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).



Environment: Basis
Mode: Steady State

Property Package Selection
- Lee-Kesler-Plocker
- MBWR
- Peng-Robinson
- PR-Twu
- PRSV
- Sour SRK
- Sour PR
- SRK
- SRK-Twu
- Twu-Sim-Tassone
- Zudkevitch-Joffee

Property Package Filter
- ○ All Types
- ● EOSs
- ○ Activity Models
- ○ Chao Seader Models
- ○ Vapour Press Models
- ○ Miscellaneous Types

EOS Enthalpy Method Specification
- ● Equation of State
- ○ Lee-Kesler

Peng Robinson Options
- ● UniSim Design
- ○ Standard

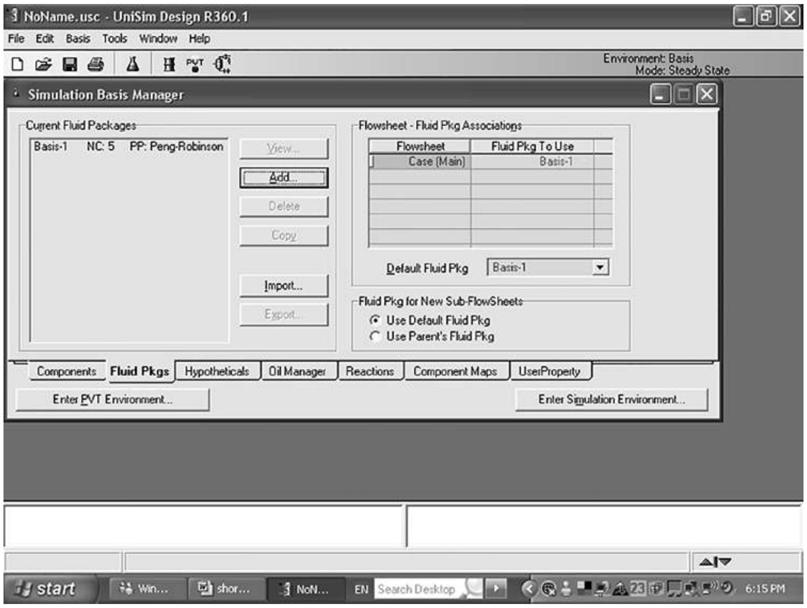
- ☐ Use EOS Density
- ☑ Smooth Liquid Density
- ☑ Modify H2 Tc and Pc
- ☐ Indexed Viscosity

Advanced Thermodynamics
- ☐ UniSim Thermo

Component List Selection: Component List - 1

Name: Basis-1
Property Pkg: Peng-Robinson

Figure 19.89 Snapshot of Simulation Basis Manager window listing Property Package Selection. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).



Environment: Basis
Mode: Steady State

Current Fluid Packages
Basis-1 NC: 5 PP: Peng-Robinson

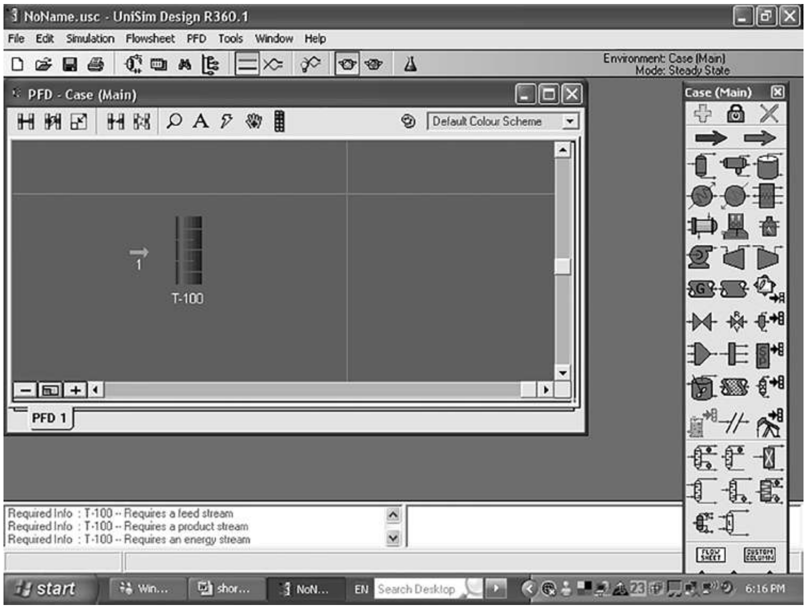
Flowsheet - Fluid Pkg Associations

Flowsheet	Fluid Pkg To Use
Case (Main)	Basis-1

Default Fluid Pkg: Basis-1

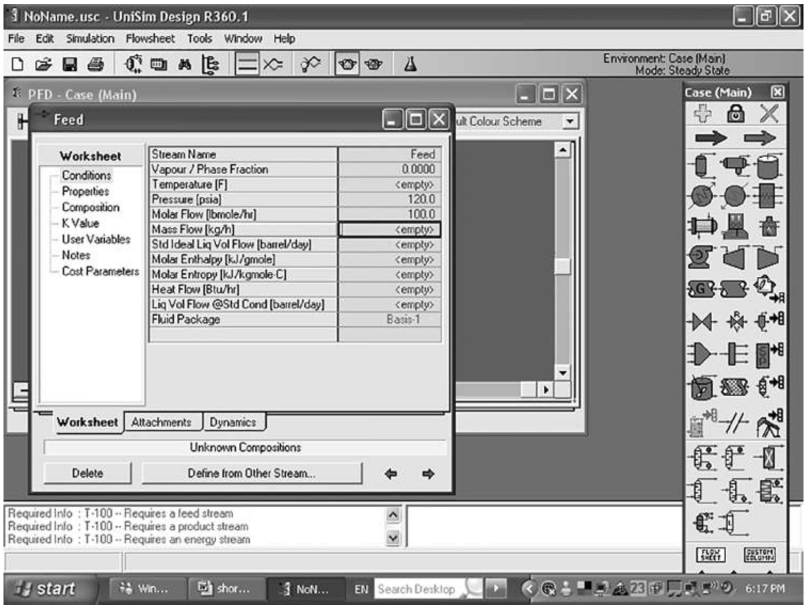
Fluid Pkg for New Sub-FlowSheets
- ● Use Default Fluid Pkg
- ○ Use Parent's Fluid Pkg

Figure 19.90 Snapshot of Simulation Basis Manager window. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).



Required Info : T-100 -- Requires a feed stream
Required Info : T-100 -- Requires a product stream
Required Info : T-100 -- Requires an energy stream

Figure 19.91 Snapshot of the feed stream 1 and distillation column T-100. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).

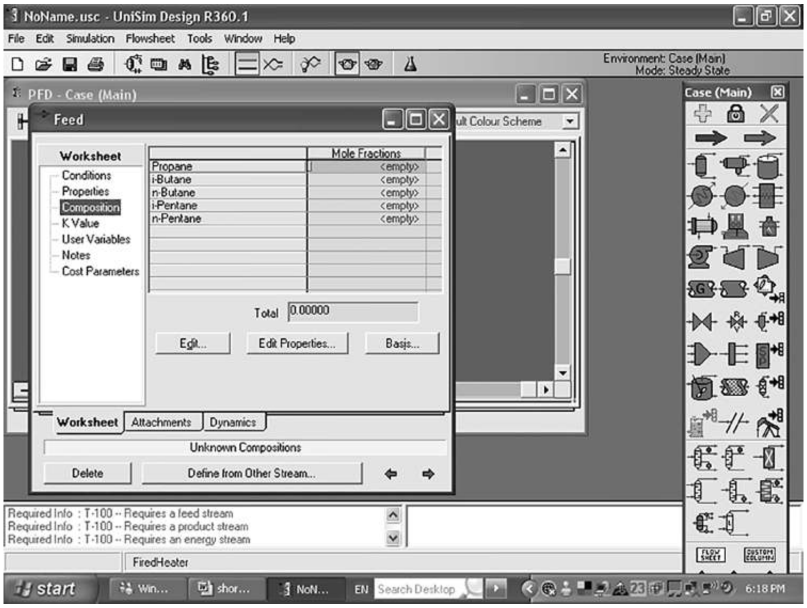


Stream Name	Feed
Vapour / Phase Fraction	0.0000
Temperature [F]	<empty>
Pressure [psia]	120.0
Molar Flow [lbmole/hr]	100.0
Mass Flow [kg/h]	<empty>
Std Ideal Liq Vol Flow [barrel/day]	<empty>
Molar Enthalpy [kJ/gmole]	<empty>
Molar Entropy [kJ/kgmole-C]	<empty>
Heat Flow [Btu/hr]	<empty>
Liq Vol Flow @Std Cond [barrel/day]	<empty>
Fluid Package	Basis-1

Unknown Compositions

Required Info : T-100 -- Requires a feed stream
Required Info : T-100 -- Requires a product stream
Required Info : T-100 -- Requires an energy stream

Figure 19.92 Snapshot of the Feed window showing the **Worksheet** parameters. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).



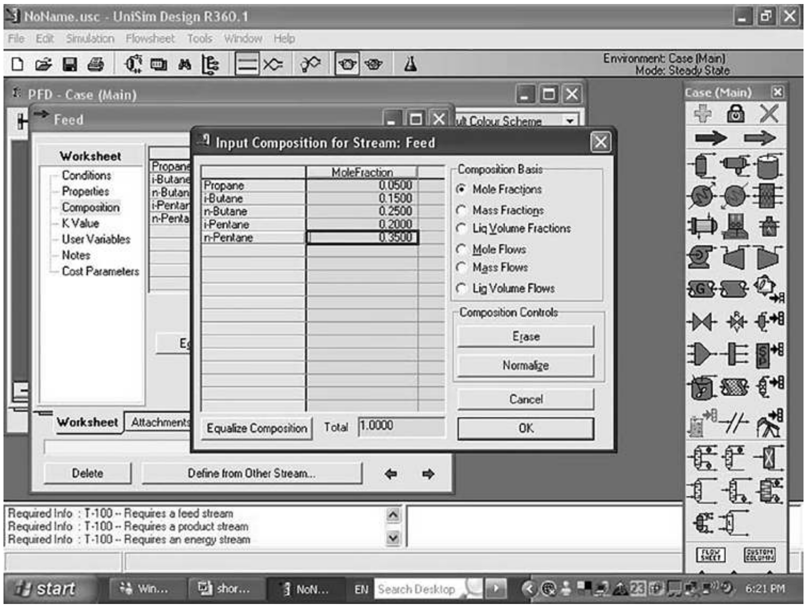
	Mole Fractions
Propane	<empty>
i-Butane	<empty>
n-Butane	<empty>
i-Pentane	<empty>
n-Pentane	<empty>

Total 0.00000

Unknown Compositions

Required Info : T-100 -- Requires a feed stream
Required Info : T-100 -- Requires a product stream
Required Info : T-100 -- Requires an energy stream

Figure 19.93 Snapshot of the Feed window with the selected five components. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).



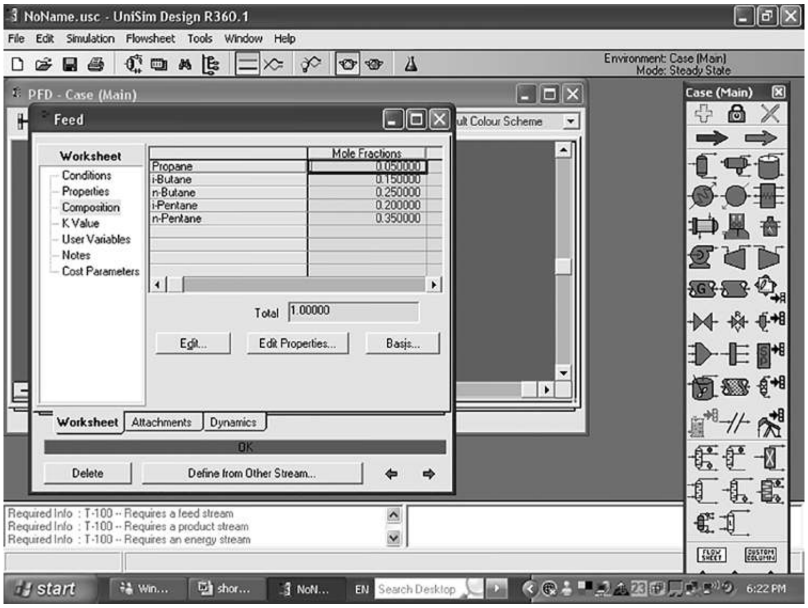
	MoleFraction
Propane	0.0500
i-Butane	0.1500
n-Butane	0.2500
i-Pentane	0.2000
n-Pentane	0.3500

Total 1.0000

Composition Basis
● Mole Fractions
○ Mass Fractions
○ Liq Volume Fractions
○ Mole Flows
○ Mass Flows
○ Liq Volume Flows

Required Info : T-100 -- Requires a feed stream
Required Info : T-100 -- Requires a product stream
Required Info : T-100 -- Requires an energy stream

Figure 19.94 Snapshot of **Input Composition** for **Stream: Feed** window. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).

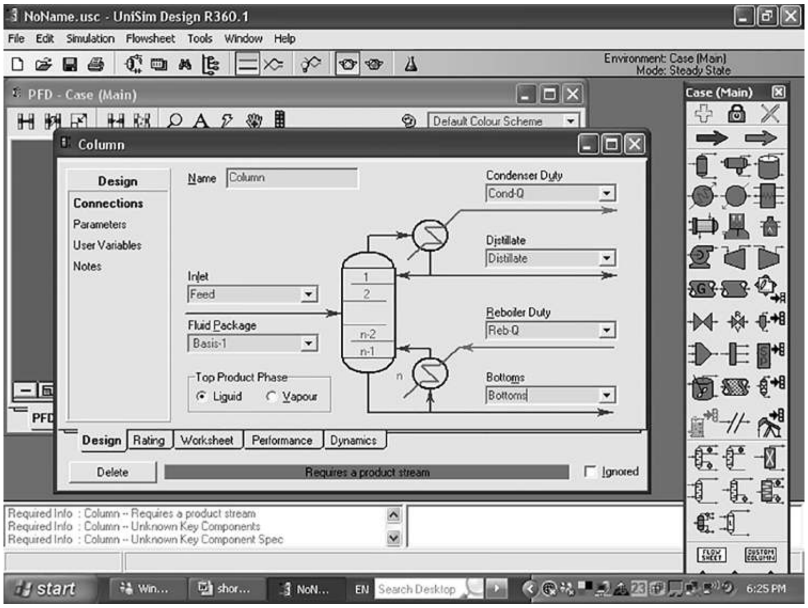


	Mole Fractions
Propane	0.050000
i-Butane	0.150000
n-Butane	0.250000
i-Pentane	0.200000
n-Pentane	0.350000

Total 1.00000

Required Info : T-100 -- Requires a feed stream
Required Info : T-100 -- Requires a product stream
Required Info : T-100 -- Requires an energy stream

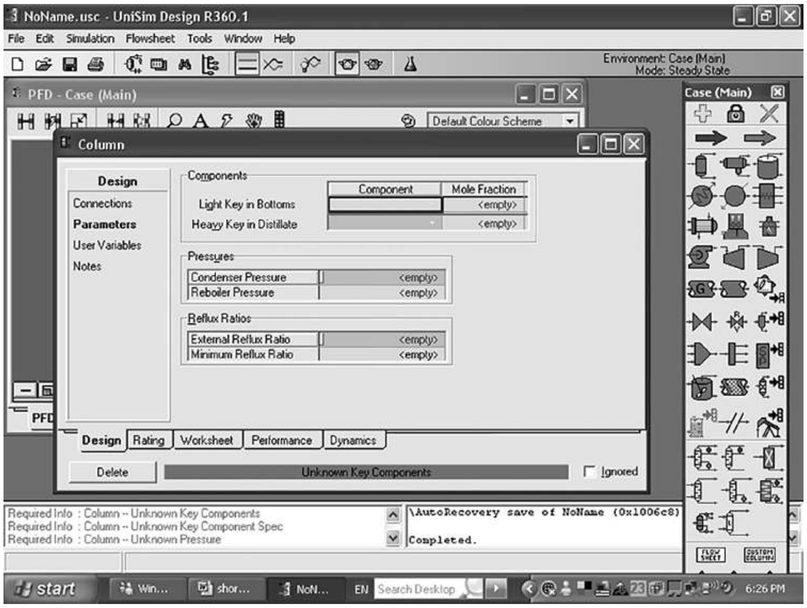
Figure 19.95 Snapshot of the Feed window showing the total mole fraction of the components. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).



Name: Column
Inlet: Feed
Fluid Package: Basis-1
Top Product Phase: ● Liquid ○ Vapour
Condenser Duty: Cond-Q
Distillate: Distillate
Reboiler Duty: Reb-Q
Bottoms: Bottoms
Requires a product stream
☐ Ignored

Required Info : Column -- Requires a product stream
Required Info : Column -- Unknown Key Components
Required Info : Column -- Unknown Key Component Spec

Figure 19.96 Snapshot of **Column** window from **Design** tab in **Connections** menu. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).



Components

	Component	Mole Fraction
Light Key in Bottoms		<empty>
Heavy Key in Distillate		<empty>

Pressures

Condenser Pressure	<empty>
Reboiler Pressure	<empty>

Reflux Ratios

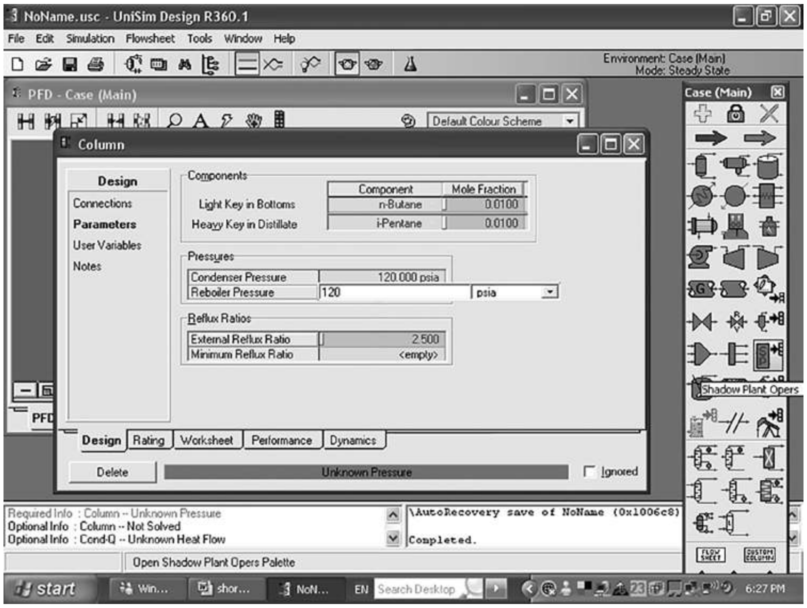
External Reflux Ratio	<empty>
Minimum Reflux Ratio	<empty>

Unknown Key Components

```
Required Info : Column -- Unknown Key Components
Required Info : Column -- Unknown Key Component Spec
Required Info : Column -- Unknown Pressure
```

```
\AutoRecovery save of NoName (0x1006c8)
Completed.
```

Figure 19.97 Snapshot of **Column** window showing **Parameters** menu. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).



Components

	Component	Mole Fraction
Light Key in Bottoms	n-Butane	0.0100
Heavy Key in Distillate	i-Pentane	0.0100

Pressures

Condenser Pressure	120.000 psia
Reboiler Pressure	120 psia

Reflux Ratios

External Reflux Ratio	2.500
Minimum Reflux Ratio	<empty>

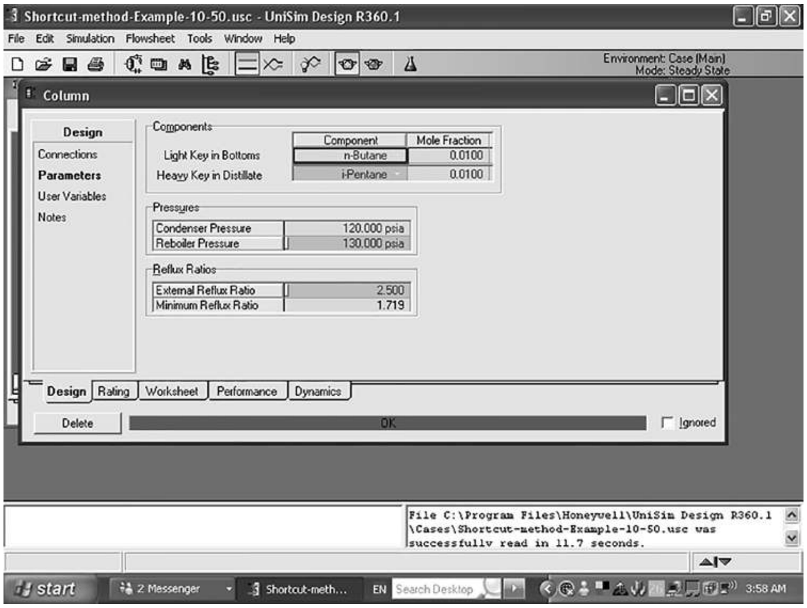
Unknown Pressure

```
Required Info : Column -- Unknown Pressure
Optional Info : Column -- Not Solved
Optional Info : Cond-Q -- Unknown Heat Flow
```

```
\AutoRecovery save of NoName (0x1006c8)
Completed.
```

Open Shadow Plant Opers Palette

Figure 19.98 Further snapshot of **Column** window showing **Parameters** menu. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).



Components

	Component	Mole Fraction
Light Key in Bottoms	n-Butane	0.0100
Heavy Key in Distillate	i-Pentane	0.0100

Pressures

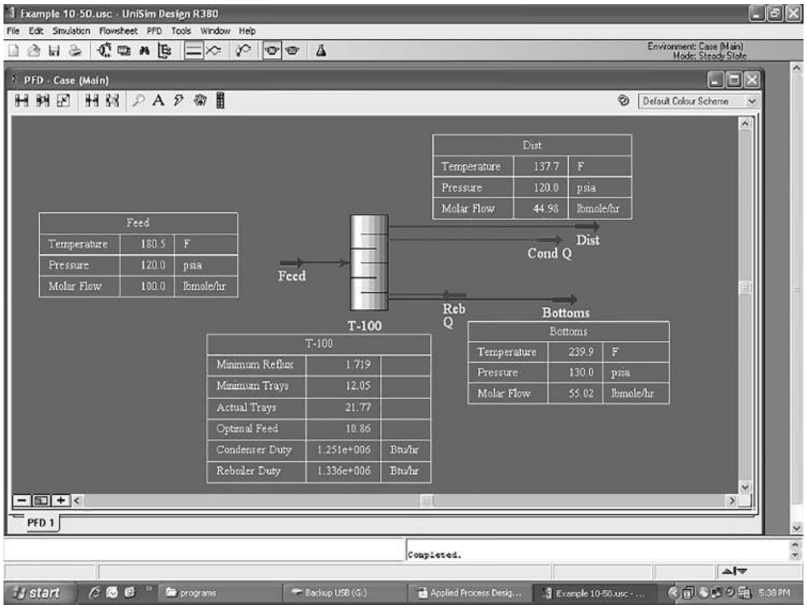
Condenser Pressure	120.000 psia
Reboiler Pressure	130.000 psia

Reflux Ratios

External Reflux Ratio	2.500
Minimum Reflux Ratio	1.719

```
File C:\Program Files\Honeywell\UniSim Design R360.1
\Cases\Shortcut-method-Example-10-50.usc was
successfully read in 11.7 seconds.
```

Figure 19.99 Snapshot of Column window with required Parameters. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).



Dist		
Temperature	137.7	F
Pressure	120.0	psia
Molar Flow	44.98	lbmole/hr

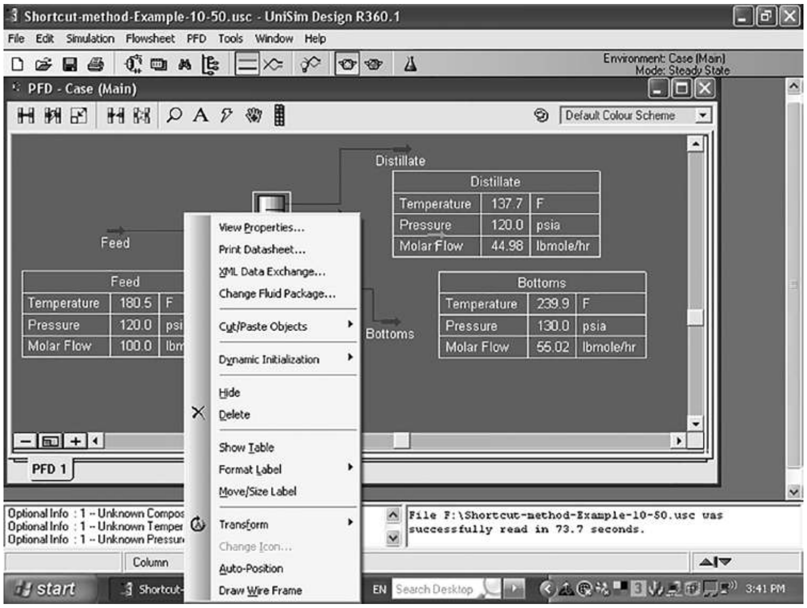
Feed		
Temperature	180.5	F
Pressure	120.0	psia
Molar Flow	100.0	lbmole/hr

T-100		
Minimum Reflux	1.719	
Minimum Trays	12.05	
Actual Trays	21.77	
Optimal Feed	10.86	
Condenser Duty	1.251e+006	Btu/hr
Reboiler Duty	1.336e+006	Btu/hr

Bottoms		
Temperature	239.9	F
Pressure	130.0	psia
Molar Flow	55.02	lbmole/hr

```
Completed.
```

Figure 19.100 Snapshot of PFD window showing the shortcut method simulation results of Example 19.42. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).



Distillate		
Temperature	137.7	F
Pressure	120.0	psia
Molar Flow	44.98	lbmole/hr

Feed		
Temperature	180.5	F
Pressure	120.0	psi
Molar Flow	100.0	lbm

Bottoms		
Temperature	239.9	F
Pressure	130.0	psia
Molar Flow	55.02	lbmole/hr

Optional Info : 1 -- Unknown Compos
Optional Info : 1 -- Unknown Temper
Optional Info : 1 -- Unknown Pressur

```
File F:\Shortcut-method-Example-10-50.usc was
successfully read in 73.7 seconds.
```

Figure 19.101 Snapshot of PFD window showing the **Print Datasheet** menu. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).

Checked Datablocks will be included in the Datasheet. Datablocks with trailing ellipses have additional options that will appear here when selected.

Feed		
Temperature	180.5	F
Pressure	120.0	psia
Molar Flow	100.0	lbm

Optional Info : 1 -- Unknown Compositions
Optional Info : 1 -- Unknown Temperature
Optional Info : 1 -- Unknown Pressure

```
File F:\Shortcut-method-Example-10-50.usc was
successfully read in 73.7 seconds.
```

Figure 19.102 Snapshot of **Select Data block(s) to Print for Shortcut Column** window. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).

Table 19.40 Multicomponent System Fractionation of five components, 99% LK recovery and 99% HK recovery of Example 19.42.

Multicomponent System Fractionation

5 components 99.0% LK recovery and 99.0% HK recovery

Component Number	**Feed Moles**	**EQUIL. CONST**	**REL. VOL.α**	**Mol. Fraction xF**	**Product in Distillate, m**	**Distillate %**	**Moles**	**xD**	**Bottoms %**	**Moles**	**xB**
1. C3H8	0.05	2.943	4.567747944	0.05	1	100	0.049999998	0.1112	4E-08	2E-09	4E-09
2. iC4H10	0.15	1.515	2.351389104	0.15	0.9994618	99.94618	0.149919265	0.3334	0.0005	8E-05	0.0001
3. nC4H10	0.25	1.232	1.912152721	0.25	0.99	99	0.2475	0.5504	0.01	0.0025	0.0045
4. iC5H12)	0.2	0.6443	1	0.2	0.01	1	0.002	0.0044	0.99	0.198	0.3598
5. nC5H12	0.35	0.5392	0.836877231	0.35	0.0008083	0.080827	0.000282894	0.0006	0.9992	0.3497	0.6355
	1			1			0.449702157	0.8888		0.5503	1

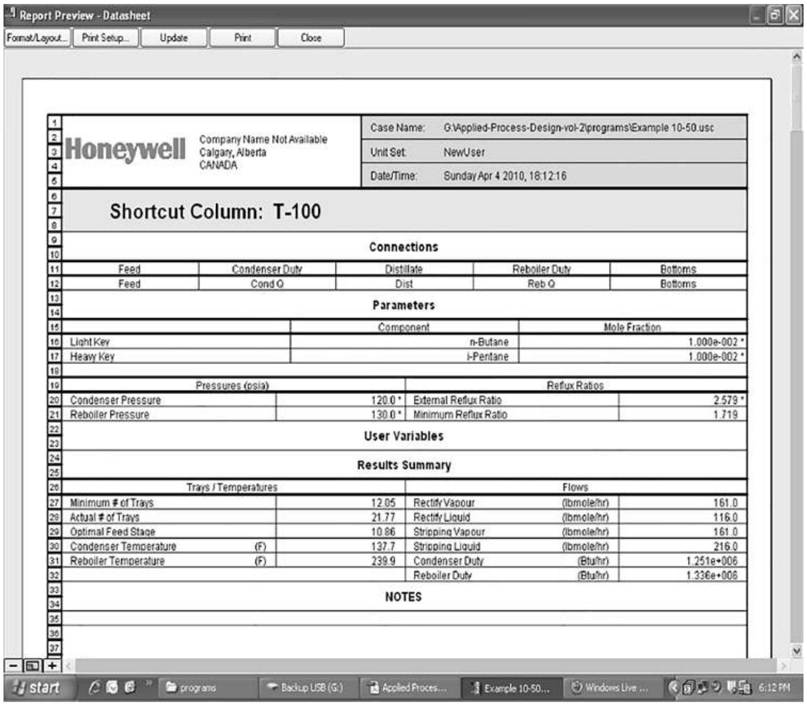
Table 19.41 Summary of the results of Example 19.42.

Parameters	Honeywell UniSim ®	Excel (Example 19.42.xlsx)
Feed molar flow rate, lbmole/hr.	100	100
Distillate molar flow rate, lbmole/hr	44.98	44.97
Bottoms molar flow rate, lbmole/hr.	55.02	55.03
Minimum reflux ratio	1.697	1.59
Actual reflux ratio	2.5	2.07
Minimum number of stages	12.5	14.2
Actual number of stages	22.34	28.4
Optimum feed stage location	11.14	14.4

The tray temperature is obtained from

$$K_h = \frac{1}{\sum \alpha_i x_i} \tag{19.451}$$

K_h is evaluated at the column pressure by use of suitable K charts.



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name: G:\Applied-Process-Design-vol-2\programs\Example 10-50.usc
Unit Set: NewUser
Date/Time: Sunday Apr 4 2010, 18:12:16

Shortcut Column: T-100

Connections

Feed	Condenser Duty	Distillate	Reboiler Duty	Bottoms
Feed	Cond Q	Dist	Reb Q	Bottoms

Parameters

	Component	Mole Fraction
Light Key	n-Butane	1.000e-002 *
Heavy Key	i-Pentane	1.000e-002 *

Pressures (psia)		Reflux Ratios	
Condenser Pressure	120.0 *	External Reflux Ratio	2.579 *
Reboiler Pressure	130.0 *	Minimum Reflux Ratio	1.719

User Variables

Results Summary

Trays / Temperatures			Flows		
Minimum # of Trays		12.05	Rectify Vapour	(lbmole/hr)	161.0
Actual # of Trays		21.77	Rectify Liquid	(lbmole/hr)	116.0
Optimal Feed Stage		10.86	Stripping Vapour	(lbmole/hr)	161.0
Condenser Temperature	(F)	137.7	Stripping Liquid	(lbmole/hr)	216.0
Reboiler Temperature	(F)	239.9	Condenser Duty	(Btu/hr)	1.251e+006
			Reboiler Duty	(Btu/hr)	1.336e+006

NOTES

Figure 19.103 Snapshot of shortcut Column simulation results of Example 19.42. (Courtesy of UniSim Design® R360.1, Honeywell Process Solutions, Honeywell® and UniSim® are registered trademarks of Honeywell International Inc.).

Stripping section: reference component is heavy key, x_h, y_h

$$\left(\frac{y_i}{y_h}\right)_m = \alpha_i\left[\frac{(V_s/B)(y_i)_{m-1} + x_{Bi}}{(V_s/B)(y_h)_{m-1} + x_{Bh}}\right] \tag{19.452}$$

$$\sum (y_i/y_h)_m = 1.0 \text{ (including } y_b\text{)}$$

The composition of each component on a tray is obtained from $(y_i/y_h)_m$.

The tray temperature is obtained from

$$K_h = \Sigma y_i/\alpha_i \tag{19.453}$$

At the column pressure using K charts for the heavy key or reference component.

Procedure

A. Rectifying Section

1. Determine material balance around column, including reflux L, distillate product D, and bottoms product B.
   a. With total condenser, the reflux composition is equal to the condensed distillate product composition.
   b. With a partial condenser, the product D is a vapor, so a dew point must be run on its composition to obtain the liquid reflux composition.
2. Determine top tray temperature for use in relative volatility calculations by running a dew point on the overhead vapor. For total condenser its composition is same as distillate product. For a partial condenser, run a dew point on the column overhead vapor composition as determined by a material balance around the partial condenser, reflux, and product.
3. Determine $(x_i/x_h)_2$, for tray No. 2 (second from top), for each component, using the x values for the reflux as the initial x_i (n + 1).
4. Total this column to yield (x_i/x_h). This equals $1/x_h$.
5. Determine x_i for each component by

$$x_i = \frac{(x_i/x_h)}{\sum (x_i/x_h)} \tag{19.454}$$

   This is liquid composition on tray.
6. Continue down column using the composition calculated for tray above to substitute in Equation (19.450) to obtain the (x_i/x_h) for the tray below.
7. Test to determine if α is varying to any great extent by calculating $\alpha_i x_i$ for a test tray. $\Sigma\alpha_i x_i = 1/K_h$. Determine temperature and evaluate corresponding values. Use new α_i if significantly different.
8. Continue step-wise calculations until the ratio of light to heavy key on a tray equals (or nearly so) that ratio in the liquid portion of the feed. This is then considered the feed tray.

9. If there are components in the feed and bottoms that do not appear in the overhead product, they must gradually be introduced into the calculations. The estimated position above the feed tray to start introducing these components is determined by

$$\frac{x_{Fi}}{x_a} = \left[\frac{1}{(1+D/L)K_i}\right]^{p''} \tag{19.455}$$

where
x_{Fi} = mol fraction of a component in the feed that does not appear in overhead
x_a = small arbitrary mol fraction in the liquid p″ plates above the feed plate
p″ = number of plates above the feed where introduction of components should begin

B. Stripping Section

1. Determine bubble point temperature of bottoms and composition of vapor, y_{Bi} up from liquid. Calculate relative volatility of light to heavy component at this temperature.
2. From these calculate vapor compositions; using Equation (19.452), calculate the ratio (y_i/y_h) for the first tray at the bottom.
3. Total $\Sigma(y_i/y_h)$ to obtain l/y_h
4. Calculate y_i for tray one

$$y_i = \frac{(y_i/y_h)}{\sum y_i/y_h}, \sum y_i/y_h = 1/y_h$$
$$\sum y_i = 1.0 \tag{19.456}$$

5. Calculate (y_i/y_h) for next tray using the y_i values of tray one $(m - 1)$ in the equation to solve for $(y_i/y_h)_m$.
6. Test to determine if α_i is varying significantly by $K_h = \Sigma(y_i/\alpha_i)$. Evaluate temperature of heavy component at the column bottoms pressure (estimated) using K charts or the equivalent. If necessary, calculate new α_i values for each component at the new temperature. Recheck every two or three trays if indicated.
7. Introduce components lighter than the light key that are not found in the bottoms in the same general manner as discussed for the rectifying section.

$$x_{Fi}/x_a = [(1+D/L)K_i]^{p'} \tag{19.457}$$

where p′ is the number of trays below the feed tray where the component, i, is introduced in an assumed amount (usually small) x_a. Then x_{Fi} is the mol fraction of the component in the feed.
8. Continue step-wise calculations until ratio of light to heavy keys in the liquid portion of the feed essentially matches the same component ratio in the liquid on one of the trays.
9. The total number of theoretical trays in the column is the sum of those obtained from the rectifying calculations, plus those of the stripping calculations, plus one for the feed tray. This does not include the reboiler or partial condenser as trays in the column.

19.43 Tray-by-Tray Calculation of a Multicomponent Mixture Using a Digital Computer

Multicomponent distillation is much more common in process plants and refineries than the simpler binary systems. Many computer programs have been developed to handle the many iterative calculations required when the system involves 3 to possibly 10 individual components with the necessary degree of accuracy. To solve a multicomponent design properly, there should be both heat and material balance at every theoretical tray throughout the calculation.

To accommodate the step-by-step, recycling and checking for convergences requires input of vapor pressure relationships (such as Wilson's, Renon's, etc.) using the previously determined constants, latent heat of vaporization data (equations) for each component (or enthalpy of liquid and vapor). Specific heat data for each component, and possibly special solubility or Henry's Law deviations, may also be required.

There are several valuable references in the literature for developing and applying a multicomponent distillation program, including Holland [173, 128, 174], Prausnitz [175, 176], Wang and Henke [129], Thurston [177], Boston and Sullivan [178], Maddox and Erbar [158], and the pseudo-K method of Maddox and Fling [179]. Convergence of the iterative trials to reach a given criterion requires careful evaluation [180]. There are several convergence techniques available—some requiring considerably less computational time than others.

Example 19.43: Tray-to-Tray Design Multicomponent Mixture

A column is to be designed to separate the feed given below into an overhead of 99.9 mol % trichloroethylene. The top of the column will operate at 10 psig. Feed temperature is 158°F.

		Overhead		**Bottoms**	
Feed	**Mole Fraction**	**Moles**	**Mole Fraction**	**Moles**	**Mole Fraction**
(A) Trichloroethylene	0.456	0.451	0.999	0.00549	0.010
(B) β Trichloroethane	0.0555	0.00045	0.001	0.05505	0.101
(C) Perchloroethylene	0.3625	…	…	0.36250	0.661
(D) Tetras (1)	0.0625	…	…	0.0625	0.114
(E) Tetras (2)	0.0635	…	…	0.0625	0.114
	1.0000	0.45145	1.000	0.54804	1.000

Note: the material balance for overhead and bottoms is based on

- **(a)** 99.9 mol % trichlorethylene in overhead
- **(b)** 1.0 mol % trichlorethylene in bottoms.
- **(c)** 1.0 mol feed total
- **(d)** Light key = trichloroethylene
  Heavy key = β trichloroethane

Determine Overhead Temperature

Because trichlorethylene is 99.9% overhead, use it only to select boiling point from vapor pressure curves at 10 psig overhead pressure = 223°F (1,280 mm Hg abs).

Determine Bottoms Temperature (Bubble Point)

Allowing 10 psig column pressure drop, bottoms pressure = 20 psig (1,800 mm Hg abs).

Component	**x_{iB}**	**Try T = 320°F Vapor Press. mm Hg**	**x_i (vp.)**	**$(yi)_B$**
A	0.01	4,500	45	0.0249
B	0.101	2,475	250	0.1382
C	0.661	1,825	1,210	0.67
D	0.114	1,600	183	0.1012
E	0.114	1,050	120	0.0664
			1,808	1.0007
			mm Hg abs.	

This compares quite well with the selected 1,800 mm bottoms pressure. Bottoms temperature is 320°F.

Relative volatilities: Light to Heavy key

$$\text{At top: } \alpha = \frac{\text{v.p. Trichloroethylene}}{\text{v.p. } \beta \text{ Trichloroethane}} = \frac{1280}{600} = 2.13$$

$$\text{At bottoms: } \alpha = \frac{\text{v.p. Trichlorethylene}}{\text{v.p. } \beta \text{ Trichloroethane}} = \frac{4500}{2275} = 1.98$$

$$\alpha \text{ (average)} = [(2.13)\ (1.98)]^{1/2} = 2.06$$

Minimum Stages at Total Reflux

$$S_M = N_{min} + 1 = \frac{\log(x_{D1}/x_{Dh})(x_{Bh}/x_{B1})}{\log \alpha_{avg}}$$

$$= \frac{\log(0.999/0.001)(0.101/0.01)}{\log 2.06}$$

$$= \frac{4.003}{0.318} = 12.6 \text{ theoretical stages}$$

Minimum Stages Above Feed

$$S_r = \frac{\log(0.999/0.001)(0.0555/0.456)}{\log 2.13} = \frac{2.082}{0.328} = 6.35 \text{ theoretical stages}$$

19.44 Thermal Condition of Feed

Feed temperature = 158°F

Calculated bubble point of feed = 266°F at assumed teed tray pressure of 15 psig.

$$q = \frac{\text{Heat to bring feed to boiling point} + \text{Heat to vaporize feed}}{\text{Latent heat of one mol of feed}}$$

$q = 1.298$ (Calculations not shown, but handled in same manner as for example given in binary section; however, all feed components are considered, not just keys.)

19.45 Minimum Reflux-Underwood Method, Determination of α_{Avg} for Multicomponent Mixture

Assume pinch temperatures (usually satisfactory because α does not vary greatly) at ⅓ and ⅔ of overall column temperature differences.

$$\text{Lower pinch} = 320 - ⅓\,(320 - 223) = 288°\text{F}$$

$$\text{Upper pinch} = 320 - ⅔\,(320 - 223) = 255°\text{F}$$

	@ 255°F		**@ 288°F**		
Component	**V.P.**	**α**	**V.P.**	**α**	**α_i (avg)**
A	2050	2.00	3050	1.91	1.955
B	1025	1.00	1600	1.00	1.00
C	750	0.732	1180	0.737	0.735
D	650	0.634	1035	0.647	0.641
E	390	0.380	650	0.406	0.393

To start, assume θ = 1.113 (it must lie between 1.00 and 1.955).

Component	**x_{Fi}**	**$\alpha_i x_{Fi}$**	**$(\alpha_i - \theta)$**	**$\alpha_i x_{Fi}/(\alpha_i - \theta)$**	**$\alpha_i x_{Fi}/(\alpha_i - \theta)^2$**
A	0.456	0.891	0.842	1.058	1.252
B	0.0555	0.0555	−0.113	−0.491	4.33
C	0.3625	0.266	−0.378	−0.704	1.86
D	0.0625	0.0401	−0.472	−0.085	0.18
E	0.0625	0.0246	−0.720	−0.0342	0.0472
				Σ = −0.2562	Σ 7.669

$$\theta_c = 1.113 - (-0.2562/7.669) = 1.113 + 0.0334$$

$$\theta_c = 1.146 \text{ (This is sufficiently close to the original to not require recalculation.)}$$

The correct value of 1.146 should be used.

Check for balance:

$$1-q=\sum\frac{x_{Fi}}{1-\theta/\alpha_i}=\sum\frac{\alpha_i x_{Fi}}{\alpha_i-\theta}=-0.256$$
$$1-1.298=-0.298=-0.256$$

This could be corrected closer if a greater accuracy were needed. This is not as good a match as ordinarily desired.

$$(L/D)_{min}+1=\frac{(\alpha_a x_a)_D}{\alpha_a-\theta}+\frac{(\alpha_b x_b)_D}{\alpha_b-\theta}$$

(for all distillate components)

$$\begin{aligned}(L/D)_{min}+1&=\frac{(1.955)(0.999)}{(1.955-1.146)}+\frac{(1.00)(0.001)}{(1.00-1.146)}\\&=2.41+(-0.00685)\\&=2.404\\(L/D)\min&=2.404-1.0=1.40\end{aligned}$$

Operating Reflux and Theoretical Trays—Gilliland Plot

Min trays = S_M = 12.6

$(L/D)_{min}$ = 1.4

Assume $(L/D)_o$	$\frac{(L/D)_o-(L/D)_M}{(L/D)_o+1}$	Read: $S-S_M/(S+1)$	Theoretical Stages S
1.4	0	∞	∞
1.6	0.0768	0.546	29
2.0	0.20	0.445	23.5
3.0	0.40	0.312	18.8
4.0	0.52	0.245	17
∞	–	–	12.6

These values are plotted in Figure 19.52. From the curve, the operating $(L/D)_o$ was selected, and the number of corresponding theoretical stages is 19. The spreadsheet Example 19.43.xlsx calculates the number of theoretical stages using the average value of the volatilities (α_{avg}) of the component at 288°F.

Tray-by-Tray Calculation—Ackers and Wade Method

Rectifying Section, $(L/D)_o$ = 3:1

Light key = Trichlorethylene; Heavy key = β Trichlorethylene

Relative Volatilities to start: Use average of top and feed

	α_{avg}
A	2.05
B	1.00
C	0.734

Neglect the heavier components than perchlorethylene components in the rectifying section.

In order to carry the perchlorethylene, it is assumed at 0.0001 mol fraction in overhead and reflux, the β-Trichloroethane is reduced to 0.0005 mol fraction for these calculations being tighter specifications than the initial calculated balance. The overall effect will be small.

Component	$x_{iD} = x_i$ **(Reflux)**	$(x_i/x_h)_1$	$(x_i)_1$	$(x_i/x_h)_2$	$(x_i)_2$
A	0.9994	975.02	0.9984	545.5	0.9971
B	0.0005	1.00	0.001024	1.0	0.001828
C	0.0001	0.273	0.000280	0.359	0.000656
		Σ = 976.293	0.999704 (sufficiently close)	546.859	

Typical calculations:

$$\left(\frac{x_i}{x_h}\right)_1 = \frac{1}{\alpha_i}\left[\frac{(L/D)(x_i)_{n+1} + x_{Di}}{(L/D)(x_h)_{n+1} + x_{Dh}}\right]$$

For component A: Tray 1

$$\left(\frac{x_i}{x_h}\right)_1 = \frac{1}{2.05}\left[\frac{(3)(0.9994)+(0.9994)}{(3)(0.0005)+0.0005}\right] = 975.02$$

Component B:

$$\left(\frac{x_i}{x_h}\right)_1 = \frac{1}{1.00}\left[\frac{3(0.0005)+0.0005}{3(0.0005)+0.0005}\right] = 1.00$$

Component C:

$$\left(\frac{x_i}{x_h}\right)_1 = \frac{1}{0.734}\left[\frac{3(0.0001)+0.0001}{3(0.0005)+0.0005}\right] = 0.272$$

$$(x_A)_1 = 975.02/976.293 = 0.9984$$

$$(x_B)_1 = 1.00/976.293 = 0.001024$$

$$(x_c)1 = 0.273/976.293 = 0.000280$$

Tray 2: Component A

$$\left(\frac{x_i}{x_h}\right)_2 = \frac{1}{2.05}\left[\frac{3(0.9984)+0.9994}{3(0.00102)+0.0005}\right] = 545.5$$

Component B

$$\left(\frac{x_i}{x_h}\right)_2 = \frac{1}{1.00}\left[\frac{3(0.00102)+0.0005}{3(0.00102)+0.0005}\right] = 1.00$$

Component C

$$\left(\frac{x_i}{x_h}\right)_2 = \frac{1}{0.734}\left[\frac{3(0.00028)+0.0001}{3(0.0005)+0.0005}\right] = 0.359$$

	$(x_i/x_h)_3$	$(x_i)_3$	$(x_i/x_h)_4$	$(x_i)_4$	$(x_i/x_h)_5$	$(x_i)_5$
A	325.24	0.9952	200.81	0.9916	126.61	0.9851
B	1.0	0.00306	1.0	0.004938	1.0	0.007781
C	0.514	0.001573	0.682	0.00337	0.908	0.007065
	326.754		202.492		128.518	

	$(x_i/x_h)_6$	$(x_i)_6$	$(x_i/x_h)_7$	$(x_i)_7$	$(x_i/x_h)_8$	$(x_i)_8$
A	80.60	0.9736	52.05	0.9520	33.97	0.9138
B	1.0	0.01208	1.0	0.01829	1.0	0.0269
C	1.213	0.01465	1.633	0.02987	2.21	0.05945
	82.813		54.683		37.18	

	$(x_i/x_h)_9$	$(x_i)_9$	$(x_i/x_h)_{10}$	$(x_i)_{10}$	$(x_i/x_h)_{11}$	$(x_i)_{11}$
A	22.47	0.8491	15.196	0.7501	7.716	0.5421
B	1.0	0.03779	1.0	0.04936	1.0	0.07026
C	2.994	0.1131	4.061	0.2005	5.516	0.3876
	26.464		20.257		14.232	

Ratio of keys in feed = 0.456/0.0555 = 8.2

Ratio of keys on Tray No. 10 = 0.7501/0.04936 = 15.2

Ratio of keys on Tray No. 11 = 0.5421/0.07026 = 7.7

Tray No. 11 should be used as feed tray (counting down from the top). Note that since the relative volatility did not change significantly between top to feed trays, the same value was satisfactory for the range.

Stripping Section

Determine V_s: per mol of feed

$$(L/V)_r = \frac{1}{1+D/L} = \frac{1}{1+\frac{1}{3}} = 0.75$$

$$V_r = \frac{(L/D)D}{(L/V)} = \frac{3(0.45145)}{0.75} = 1.806$$

$$L_r = (L/D)\,(D) = 3(0.45145) = 1.35 \text{ mols/mol feed}$$

$$L_s = L_r + qF = 1.35 + 1.298\,(1.0) = 2.648$$

$$V_s = V_r - F(1 - q) = 1.806 - (1.0)\,(1 - 1.298) = 2.104$$

$$V_s/B = 2.104/0.54804 = 3.84$$

Relative volatilities, α_i, are determined at average temperature between bottom and feed of column. Usually the pinch temperature gives just as satisfactory results.

Component	x_{iB}	y_{iB}	(α_i)avg	$(y_i/y_h)_1$	$(y_i)_1$	$(y_i/y_h)_2$	$(y_i)_2$
A	0.010	0.0249	1.905	0.319	0.0543	0.552	0.107
B	0.101	0.1382	1.00	1.000	0.170	1.00	0.194
C	0.660	0.6700	0.740	3.800	0.647	3.08	0.597
D	0.114	0.1012	0.648	0.517	0.088	0.389	0.0754
E	0.114	0.0664	0.411	0.241	0.0411	0.1476	0.0286
Total				5.877		5.1686	

Typical calculations: starting at bottom and working up the column.

Tray 1: Component A

$$(y_i/y_h)_1 = \alpha_i\left[\frac{(V_s/B)(y_i)_{m-1} + x_{iB}}{(V_s/B)(y_h)_{m-1} + x_{Bh}}\right] = 1.905\left[\frac{(3.84)(0.0249) + 0.010}{3.84(0.1382) + 0.101}\right]$$

$$(y_i/y_h)_1 = 0.319$$

$$(y_i)_1 = 0.319/5.877 = 0.0543$$

Tray 2: Component A

$$(y_i/y_h)_2 = 1.905\left[\frac{(3.84)(0.0543) + 0.010}{3.84(0.170) + 0.101}\right] = 0.552$$

Continuation of the calculations gives an approximate match of ratio of keys on plate 10 to those in feed. Then feed tray is number 10 from bottom, and this is also number 11 from top.

Liquid mol fraction ratio from vapor mol fraction ratio:

$$(x_i/x_h) = \frac{(y_i/y_h)}{\alpha_{1/h}}$$

Ratio on tray no. 9 = 15.018/1.905 = (x_i/x_h) = 7.9

Ratio on tray no. 10 = 19.1611.905 = 10.05

Ratio in feed = 8.2

Total theoretical trays = 11 + 10 – 1 (common feed tray count)

= 20 not including reboiler

Total theoretical stages = 20 + 1 (reboiler) = 21

This compares with 19 theoretical stages from Gilliland Plot.

Tray Efficiency

Use average column temperature of 271°F and feed analysis.

Component	**x_{iF}**	**μ, cP**	**μ, $x_{iF,}$ cP**	**V.P., mm Hg**	**$\alpha_{1/h}$**
A	0.456	0.28	0.128	2500	1.94
B	0.0555	0.36	0.020	1290	
C	0.362	0.37	0.134		
D	0.0625	0.40	0.025		
E	0.0625	0.48	0.030		
			Σ = 0.337 cp		
		αΣ(μ) (x_{iF}) = 1.94 (0.337) = 0.654			

Using Figure 19.54

Drickamer and Bradford curve,

$$E_o = 46\%$$

O'Connell curve,

$$E_o = 53.8\%$$

In this case, recommend using:

$$E_o = (46 + 53.8)/2 = 49.6\%$$

Actual trays in column:

$$N_{act} = 20/0.496 = 40.3 \text{ trays}$$

From tray-by-tray calculations, feed tray is 10/0.496 = 20.1 trays from bottom; use 20.

Generally, in practice, a column would be selected with extra trays, as a contingency, e.g., a column with a total = 45 trays.

	Number of trays
Rectifying trays	22
Feed	1
Stripping	22
Total	45

Feed nozzles should be located on tray Nos. 21, 23, and 25 counting up from the bottom tray as No. 1.

The Excel spreadsheet program (Example 19.43a.xlsx) has been developed to determine the minimum number of plates, Underwood's constant (θ), minimum reflux ratio, actual number of theoretical plates, and the feed location using 99.9% LK recovery and 99.9% HK recovery with β-Trichloroethane as the heavy key component. Table 19.42 shows the results of the spreadsheet calculations.

19.46 Heat Balance-Adjacent Key Systems with Sharp Separations, Constant Molal Overflow

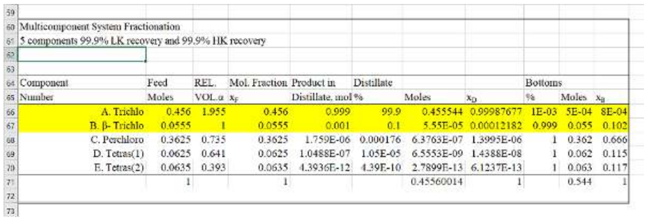
Total Condenser Duty

Refer to Figure 19.104 (System (1))

1. Assume or set condenser liquid product temperature, t_D.
2. Calculate condensing pressure, with t_D as bubble point (if subcooling exists, and t_D is below bubble point, use bubble point temperature for pressure calculation only).
3. $V_1 = L + D$

$$H_1 V_1 = [L\, h_D + D\, h_D\,] + Q_c \quad (19.458)$$

Table 19.42 Excel spreadsheet calculations of 5 components 99.9% LK recovery and 99.9% HK recovery of Multicomponent System Fractionation of Example 19.43.



Multicomponent System Fractionation

5 components 99.9% LK recovery and 99.9% HK recovery

Component Number	Feed Moles	REL. VOL. α	Mol. Fraction x_F	Product in Distillate, mol	Distillate %	Moles	x_D	Bottoms %	Moles	x_B
A. Trichlo	0.456	1.955	0.456	0.999	99.9	0.455544	0.99987677	1E-03	5E-04	8E-04
B. β- Trichlo	0.0555	1	0.0555	0.001	0.1	5.55E-05	0.00012182	0.999	0.055	0.102
C. Perchloro	0.3625	0.735	0.3625	1.759E-06	0.000176	6.3763E-07	1.3995E-06	1	0.362	0.666
D. Tetras(1)	0.0625	0.641	0.0625	1.0488E-07	1.05E-05	6.5553E-09	1.4388E-08	1	0.062	0.115
E. Tetras(2)	0.0635	0.393	0.0635	4.3936E-12	4.39E-10	2.7899E-13	6.1237E-13	1	0.063	0.117
	1		1			0.45560014	1		0.544	1

The feed condition, Q	1.0
The heavy key component number	2
Percentage recovery of the light key component in the distillate (%)	99.9
Percentage recovery of the heavy key component in the bottoms (%)	99.9
Total moles in the distillate, moles	0.4556
Total moles in the bottoms, moles	0.544
Total moles in the feed, moles	1.0
Underwood constant (θ)	1.153
Factor for the reflux ratio	1.3
Minimum reflux ratio	1.436
Actual reflux ratio	1.87
Minimum number of stages	20.61
Number of theoretical plates in the column	41.72
The position of the feed plate	25

$$H_1 = \sum_{1}^{1} H_{1i} y_{1i} \text{ at } t_1 \tag{19.459}$$

$$h_D = \sum_{1}^{i} h_{Di} x_{Di} \tag{19.460}$$

$$Q_c = V_1 (H_1 - h_D) \tag{19.461}$$

4. Calculate t_1 and x_1 by dew point on vapor V_1. Then determine H_1, referring to top tray as number one in this case, where
   H_1 = total vapor enthalpy above reference datum for sum of all contributing percentages of individual components, i, in stream. Btu/lb, or Btu/mol.

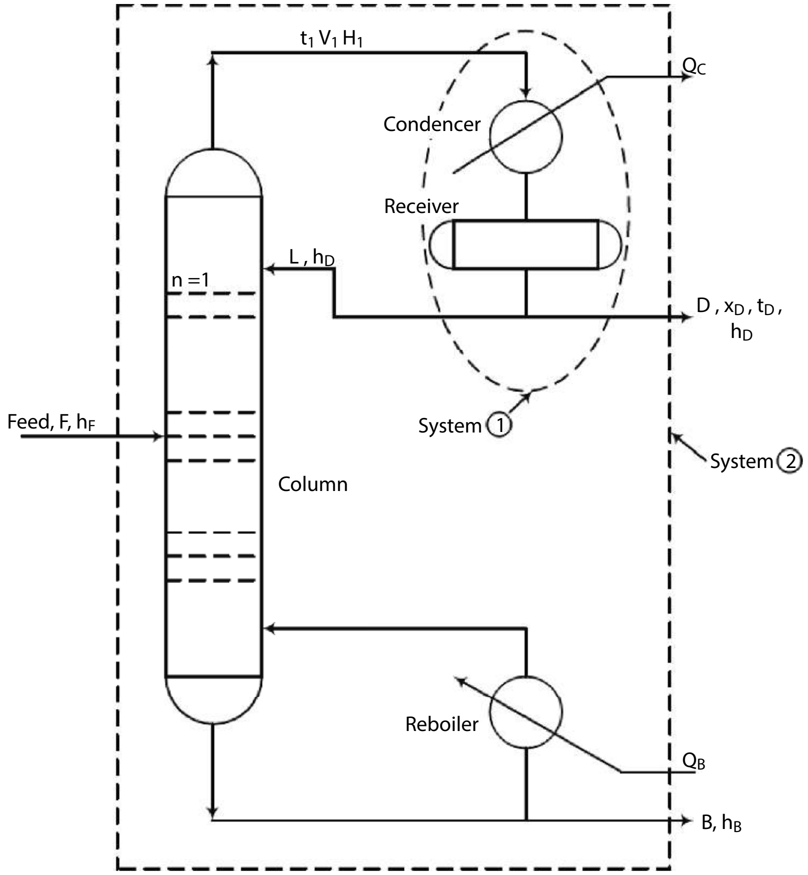


Figure 19.104 Heat balance diagram.

h_D = total liquid enthalpy above reference datum for sum of all contributing percentages of individual components, i, in product stream (also same as reflux), Btu/lb or Btu/mol.

5. For partial condenser: replace Dh_D by DH_D in Step 3. A dew point calculation on compositions of y_D (vapor) gives t_D or total pressure. Also get liquid composition x_D (liquid reflux in equilibrium with product vapor y_D). Overhead vapor is sum of compositions of y_D and x_p. A dew point calculation on this vapor (overhead from tray one top) gives top tray temperature, t_1.

$$V_1H_1 + Q_c = Lh_D + DH_D \tag{19.462}$$

Reboiler Duty

Refer to Figure 19.53 (System (2))

1. Determine bottoms temperature by bubble point on liquid x_B.
2. From feed condition determine enthalpy.

$$h_F = \frac{\sum V_F(H_i y_i)_F + \sum L_F(h_i x_i)_F}{F} \tag{19.463}$$

3. Solve for Q_B, reboiler duty, Btu/h

$$F\,h_F + Q_B = Dh_D + Bh_B + Q_c \qquad (19.464)$$

where
h_D = total enthalpy of distillate product, Btu/mol or Btu/lb
h_B = total enthalpy of bottoms product, Btu/mol or Btu/lb
h_F = total enthalpy of feed, But/mol or Btu/lb

19.47 Stripping Volatile Organic Chemicals (VOC) from Water with Air

Li and Hsiao [181] provide a useful approach to the environmental problem of removing of stripping volatile organics from solution in a contaminated water stream by using fresh air as the stripping medium, in order to reduce atmospheric pollution. A number of industrial firms perform this stripping with steam. The mass balance on the VOC component around the column (trayed or packed) as shown in Figure 19.105 uses the symbols of Reference 181.

$$\frac{L}{V} = \frac{y_i - y_{n+1}}{x_o - x_N} \text{ (slope of operating line)} \qquad (19.465)$$

where
x_o = VOC mol fraction (ratio of number of mols of a specific VOC component in water solution to the total moles of all contaminants contained in the water)

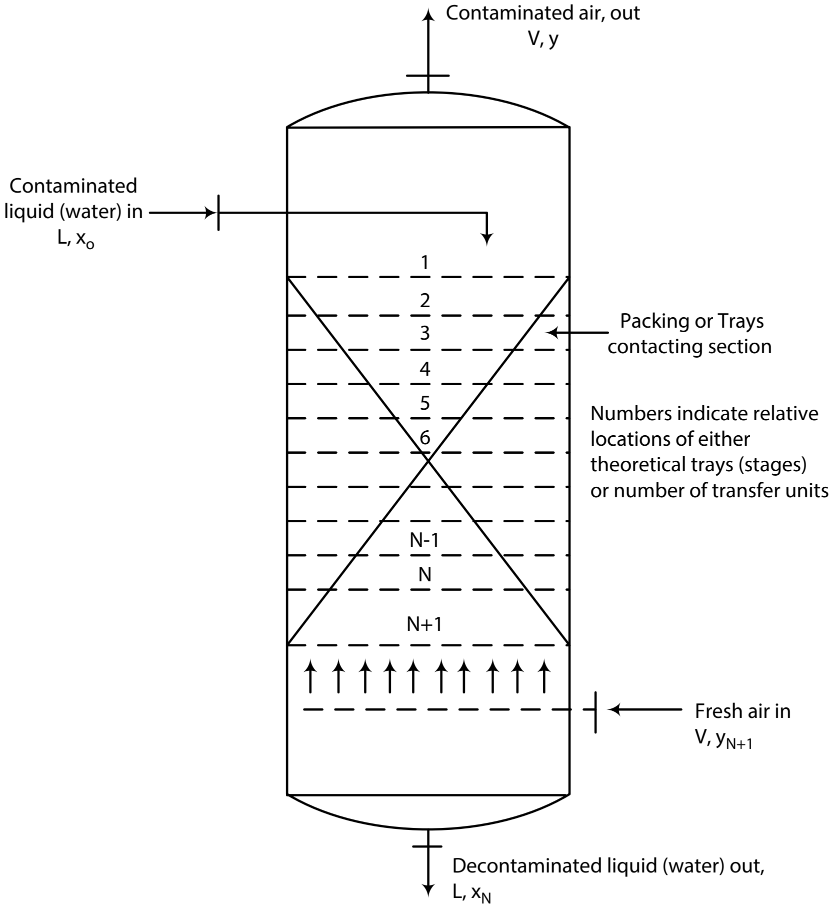


Figure 19.105 Schematic stripping tower using air to strip organics from water solution. Adapted and used by permission, Li, K. Y., and Hsiao, K. J., Chem. Eng., V. 98, No. 7, p. 114 (1991).

x_N = mol fraction of VOC component in the stripped water
N = number of theoretical trays
y_i = mol fraction VOC component in exiting VOC contaminated air
y_{N+1} = mol fraction VOC component in the incoming fresh air, = 0 for fresh air
L = volumetric flow rate for incoming contaminated water
V = volumetric flow rate for incoming fresh air
V_{min} = minimum fresh air flow required based on slope of operating line L/V on x–y diagram
x_N = mol fraction VOC contaminant in exiting water stream, usually aimed at meeting the environmental regulations
S_{min} = minimum stripping factor at minimum flow rate for air
S_{opt} = optimum stripping factor, where treatment costs are a minimum, referenced to costs of utilities, maintenance, depreciation, labor; as economic conditions change, one may need to adjust S_{opt}; see Reference 181.

The concentrations of most of the VOC compounds in the contaminated water are usually expressed in ppm as are the other residual compounds in the water exiting the tower. These are usually small values. As an approximation:

$$(L/V_{min}) = K$$

where

K = equilibrium constant (varies for each component)
K = y*/x*
y* = equilibrium molar fraction of VOC components in air
x* = equilibrium molar fractions of VOC components in water

Minimum stripping factor at corresponding minimum air flow rate:

$$\begin{aligned} S_{min} &= K/(L/V_{min}) = 1.0 \\ V_{min} &= L/K \end{aligned} \tag{19.466}$$

The component with the lowest equilibrium constant is called the key component in the stripping process, because it yields the largest value of V_{min}. This largest value is the "true" minimum air flow rate, whereas the actual air flow rate should be selected at 1.20 to 2.0 times the minimum. This is based upon an economic balance between the number of actual trays and air flow rate, which determines the column diameter.

It is relevant to evaluate the optimum stripping factor based on related costs; thus:

$$S_{opt} = K(L/V_{opt}) \tag{19.467}$$

$$V_{opt} = S_{opt}\ (V_{min}) \tag{19.468}$$

The Henry's Law constant, H, can be substituted for the equilibrium constant, K, when the system operates at or very close to atmospheric pressure:

$$H = p^*/x^* \tag{19.469}$$

where

p* = the partial pressure, atm, of the contaminant in equilibrium with x*

Tables 19.43 and 19.44 provide values for selected Henry's Law constants, respectively [181].

The optimum stripping factor, S_{opt}, is expressed as a percent of residue, (100) (x_N/x_o), for water rates of 30, 300, and 3,000 gpm.

$$S_{opt} = 1 + {}_aH_b \tag{19.470}$$

Table 19.43 Henry's Law constants and optimum stripping factors for selected organic compounds for use with Packed Towers @ 25°C (77°F).

Chemicals	Henry's Law constant	x_N/x_o %	L = 30 gpm	L = 300 gpm	L = 3,000 gpm
1,1,2,2 -tetrachloroethane	24.02	4.75	1.39	1.66	1.84
		0.05	1.88	2.30	2.59
1,1,2-trichloroethane	47.0	47.5	1.45	1.89	2.32
		0.05	2.00	2.79	3.37
1,2-dichloroethane	61.2	4.75	1.46	1.97	2.54
		0.05	2.03	2.95	3.73
Propylene dichloride	156.8	4.75	1.6	2.43	3.9
		0.05	2.3	3.9	6.13
Methylene chloride	177.4	4.75	1.57	2.37	3.90
		0.05	2.23	3.87	6.20
Chloroform	188.5	4.75	1.59	2.46	4.10
		0.05	2.28	4.05	6.61
1,1,1-trichloroethane	273.56	4.75	1.67	2.7	5.08
		0.05	2.43	4.62	8.37
1,2-dichloroethane	295.8	4.75	1.65	2.68	5.08
		0.05	2.40	4.50	8.40
1,1-dichloroethane	303.0	4.75	1.67	2.72	5.20
		0.05	2.40	4.63	8.66
Hexachloroethane	547.7	4.75	1.85	3.27	7.74
		0.05	2.7	6.0	13.6
Hexachlorobutadiene	572.7	4.75	1.88	3.48	8.1
		0.05	2.78	6.20	14.27
Trichloroethylene	651.0	4.75	1.82	3.27	7.78
		0.05	2.68	5.87	14.0
1,1-dichloroethene	834.03	4.75	1.84	3.37	8.50
		0.05	2.70	6.10	15.9
Perchloroethane	1,596.0	4.75	2.10	4.20	13.2
		0.05	3.10	7.90	26.1
Carbon tetrachloride	1,679.17	4.75	2.06	4.2	13.2
		0.05	3.1	7.9	26.45

(Used by permission, *Chem. Eng.*, Li, K. Y. and Hsiao, K. J., V. 98, No. 7 (1991), p. 114; all rights reserved).

Table 19.44 Henry's Law constants and optimum stripping factors for selected organic compounds for use with Tray Towers @ 25°C (77°F).

Chemicals	**Henry's Law constant**	x_N/x_o **%**	**L = 30 gpm**	**L = 300 gpm**	**L = 3,000 gpm**
1,1,2,2-tetrachloroethane	24.02	4.75	1.7	1.65	1.35
		0.05	2.3	2.1	1.6
1,1,2-trichloroethane	47.0	4.75	1.98	1.98	1.56
		0.05	3.10	3.00	2.18
1,2-dichloroethane	61.2	4.75	2.09	2.14	1.67
		0.05	3.34	3.31	2.40
Propylene dichloride	156.8	4.75	3.1	3.3	2.5
		0.05	5.1	5.5	4.0
Methylene chloride	177.4	4.75	2.96	3.38	2.57
		0.05	5.38	5.84	4.1
Chloroform	188.5	4.75	3.15	3.6	2.72
1,1,1-trichloroethane	273.56	4.75	4.1	4.6	3.2
		0.05	7.1	7.2	5.5
1,2-dichloroethane	295.8	4.75	3.7	4.7	3.4
		0.05	7.59	8.6	5.84
1,1-dichloroethane	303.0	4.75	3.81	4.84	3.5
		0.05	7.8	8.9	6.0
Hexachloroethane	547.7	4.75	6.3	7.5	5.1
		0.05	10.5	14.5	8.4
Hexachlorobutadiene	572.7	4.75	6.5	7.8	5.3
		0.05	11.0	15.2	9.1
Trichloroethylene	651.0	4.75	6.6	8.3	5.8
		0.05	13.9	16.9	10.9
1,1-dichloroethene	834.03	4.75	7.2	10.5	6.9
		0.05	12.0	19.2	12.0
Perchloroethane	1,596.0	4.75	9.8	13.3	11.8
		0.05	16.1	36.1	22.4
Carbon tetrachloride	1,679.17	4.75	9.2	11.7	12.1
		0.05	15.4	33.0	21.0

(Used by permission, *Chem. Eng.*, Li, K. Y. and Hsiao, K. J., V. 98, No. 7 (1991), p. 114; all rights reserved).

Constants a and b were determined from a linear regression for x_N/x_o = 4.75% and x_N and x_o = 0.05% for the packed and tray towers. The optimum stripping factor decreases as the Henry's Law constant decreases. Due to the complex relationship between cost and performance, the authors [181] recommend caution in attempting to extrapolate from the water flow rate ranges shown.

Example 19.44: Stripping Dissolved Organics from Water in a Packed Tower Using Method of Li and Hsiao [181]

Using a packed tower, remove hexachloroethane (HCE) at a concentration of 110 ppm in water to 0.05 ppm using fresh air operating at essentially atmospheric pressure using a fan/blower putting up 1½ in. water pressure. The concentration of propylene dichloride (PDC) in the contaminated water is 90 ppm and is to be reduced to 0.05 ppm in the exiting water. The water flow rate = 300 gpm. The required packing (or trays) must be determined by using a vapor–liquid equilibrium plot, setting slope L/V and stepping off the number of stages or transfer units. See Figure 19.105.

From Table 19.43 (Packed Tower):

Hexachloroethane: Henry's Law constant = 547.7 atm
Propylene dichloride: Henry's Law constant = 156.8 atm

1. For Hexachloroethane: x_N/x_o = 0.05 ppm/100 ppm = 0.05%
   For propylene dichloride: x_N/x_o = 0.05 ppm/100 ppm = 0.05%
2. S_{opt} = 6.0 for HCE, and 3.9 for PDC.
3. For HCE:

$$\begin{aligned} V_{min} &= L/K = (300)\ (8.33)\ (359 \text{ scf/mol})/(18\text{lb/mol})\ (547.7) \\ &= 91.1 \text{ scf/min.} \end{aligned}$$

For PDC:

$$\begin{aligned} V_{min} &= L/K= (300 \text{ gpm})\ (8.33)\ (359)/(18)\ (156.8) \\ &= 317.8 \text{ scf/min} \end{aligned}$$

4. Use the larger air rate as control required, which is 317.8 scf/min required for PDC, to calculate the optimum flow rate.
5. V_{opt} = (3.9) (317.8) = 1,239.4 scf/min

$$S_{opt} = K/(L/V_{opt})$$

$$V_{opt} = S_{opt}\ (V_{min})$$

6. Therefore, the operating conditions would be

$$L = 300 \text{ gpm}$$

   V = 1,239 scf/min (minimum, may want to consider actually using 10–15% more for some assurance that the required conditions will be met).
7. Determine the tower diameter based on the flows of (6) above. See Chapter 20, this volume, for packed tower design.

19.48 Rigorous Plate-to-Plate Calculation (Sorel Method [182])

The plate-to-plate calculation by Sorel using mole fractions is outlined as follows:

1. **a.** For a total condenser as shown in Figure 19.106, the temperature of plate 1 is calculated as the dew point temperature of the vapor having the same composition as the distillate D. Therefore, a trial-and-error calculation is carried out until the relation

$$\sum x_i = 1.0 = \sum \frac{y_i}{K_i} = \sum \frac{y_i \phi_i P_t}{\gamma_i P_i} \tag{19.471}$$

is satisfied. This is carried out by selecting a number of temperatures (usually a minimum of three) and obtaining K or equivalent values at the column pressure and the selected temperature for each component, and then summing the y/K values. This is followed by plotting the summation versus temperature and reading the temperature where the summation equals 1.0, as shown in Figure 19.107. The liquid composition in equilibrium with the vapor from plate 1 is calculated by obtaining the K values at the temperature and pressure on plate 1 and by evaluating y/K for each component.

b. For a partial condenser as shown in Figure 19.108, the temperature of the vapor product and reflux liquid must be determined. This temperature is known as the dew-point of the vapor and is calculated in the same manner as described in part 1a. A plot of the summation versus temperature allows the temperature corresponding to the point where the sum equals unity to be read. The composition of the reflux liquid L_o is determined by obtaining K values for each component at the calculated temperature and pressure, and determining $x_i = y_i/K$.

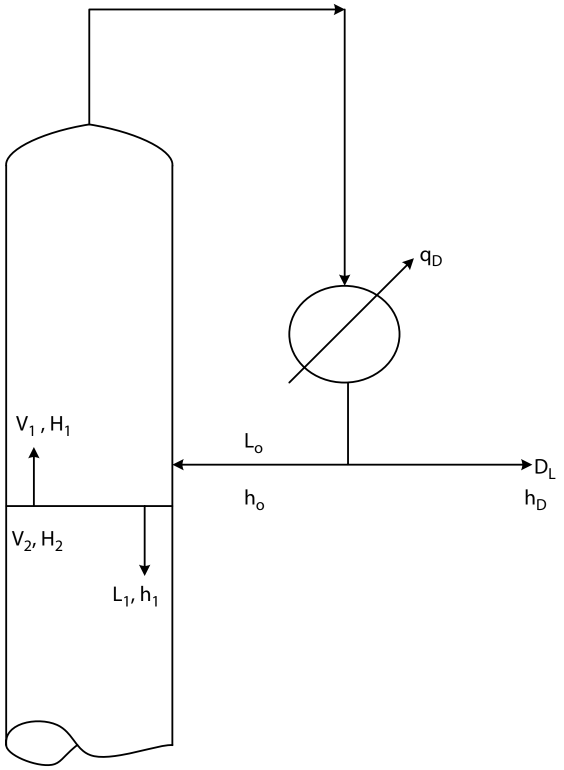


Figure 19.106 Total condenser.

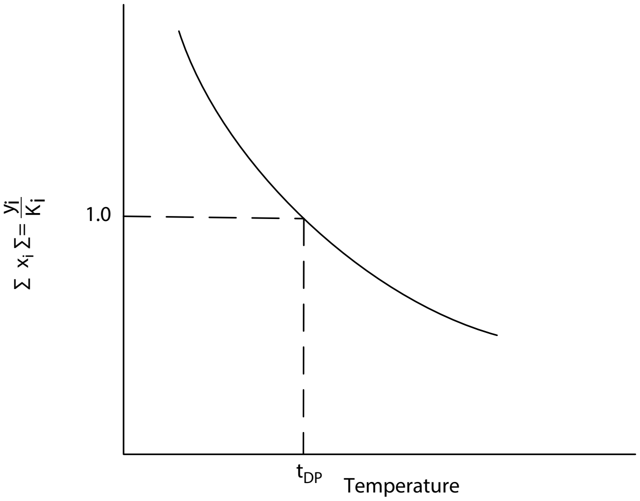


Figure 19.107 Determination of dew point.

2. The quantity of vapor rising from plate 1 is determined by material balance. Since L_o/D and D are known or fixed:

$$V_1 = L_o + D \tag{19.472}$$

or

$$V_1 = \left(\frac{L_o}{D} + 1\right)D \tag{19.473}$$

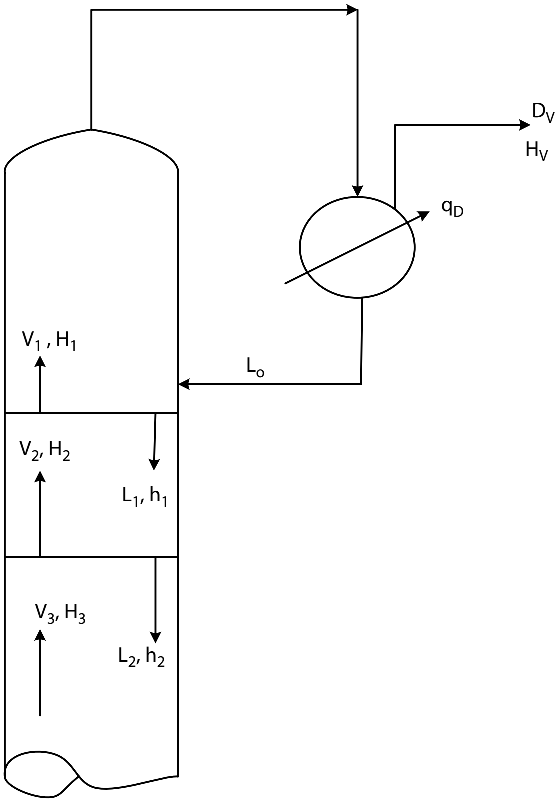


Figure 19.108 Partial condenser.

3. The condenser duty or heat removed in the condenser is then calculated by enthalpy balance, since all temperatures and pressures are known.

$$q_D = L_o h_o + D h_D - V_1 H_1 \tag{19.474}$$

4. The composition of the vapor V_2 rising to plate 1 (or V_1 in the case of partial condenser) is calculated by material and enthalpy balance. The equations for each case are illustrated as follows:

Total Condenser	**Partial Condenser**	
$V_2 = L_1 + D$	$V_1 = L_o + D$	(19.475)
$V_2 y_{i2} = L_1 x_{il} + D x_{iD}$	$V_1 y_{il} = L_o x_{io} + D x_{iD}$	(19.476)
$V_2 H_2 + q_D = L_1 h_1 + D h_D$	$V_1 H_1 + q_D = L_o h_o + D h_D$	(19.477)
$H_2 = f(y_{i2}, y_{j2}, y_{k2}, \ldots\ldots\ldots T_2, P_t)$	$H_1 = f(y_{il}, y_{jl}, y_{kl}, \ldots\ldots\ldots T_1, P_t)$	(19.478)

With a total condenser and reflux liquid cooled to a temperature below its bubble point, as shown in Figure 19.109, there are insufficient equations to determine the composition of V_2, quantity of V_2 and L_1, and the temperature T_2. Assuming that the molal latent heat of the mixture is essentially constant and varies very little with temperature and composition, and that the molal heat capacities of the components in the system are essentially constant, and vary only slightly with temperature, then the following equations can be used to determine the solution:

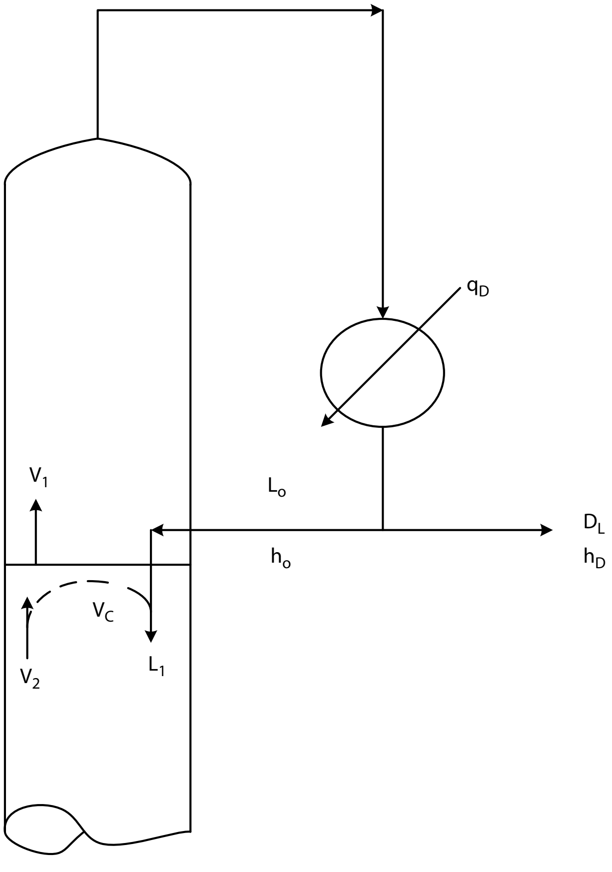


Figure 19.109 Reflux caused by supercooled liquid reflux from condenser.

$$V_C\lambda_1 = L_o\,C_{PL}\,(T_o - T_1) \tag{19.479}$$

$$V_C + V_1 = V_2 \qquad L_o + V_C = L_1 \tag{19.480}$$

$$T_2 = \frac{DC_{P_L}(T_o - T_1) - q_D - V_2\lambda_1}{V_2\,C_{P_{V1}}} + T_1 \tag{19.481}$$

5. With known V_2, the composition can be determined by Equation (19.476), since the quantities L_1 and D and the compositions x_{i1} and x_{iD} are known.

   Note: Enthalpies of the streams may be calculated approximately by the following:

$$h_o = \sum_1^n \bar{h}_o$$
$$= x_{io}\,C_{pi}(T_o - T_r) + x_{jo}\,C_{p_j}(T_o - T_r) + \ldots\ldots\ldots + x_{no}C_{p_n}(T_o - T_r)$$
$$h_o = h_D \text{ with a total condenser}$$

   For a partial condenser:

$$H_D = \sum_1^n \bar{H}_D = \sum_1^n x_i\left[\lambda_{i_r} + C_{P_v}(T_D - T_r)\right]$$

6. Determine the composition of L_2 in equilibrium with V_2 at T_2 and the column pressure by calculating the equilibrium concentrations x_{i2} by

$$x_{i2} = \frac{y_{i2}}{K_{i2}} \tag{19.482}$$

7. The vapor V_2 rising to plate 2 can be determined if the partial molal enthalpies of the components in the mixture are known from equations comparable to Equations (19.475) through (19.478). If the Lewis–Matheson assumptions of equimolal overflow and boil-up are made, V_3 equals V_2.
8. Determine the composition of V_3 by component and material balance.
9. Calculate the temperature of V_3 by the equation comparable to Equation (19.481).
10. Repeat the calculations plate by plate until a vapor composition is reached that approximates the composition of the vapor portion of the feed, or until a liquid composition of the vapor portion of the feed, or until a liquid composition is reached that approximates the composition of the liquid portion of the feed.

    Note: As a guide, if the calculated ratio of the compositions of the key components approximately matches that of the ratio of keys in the feed, and if the calculated temperature essentially matches the temperature of the feed, the number of equilibrium stages is approximately correct.

11. From Figure 19.110, the quantity of liquid and vapor below the feed plate can be determined by

$$\bar{L}_F = L_{F-1} + F \tag{19.483}$$

$$\bar{V}_{F+1} = V_F - F_V \tag{19.484}$$

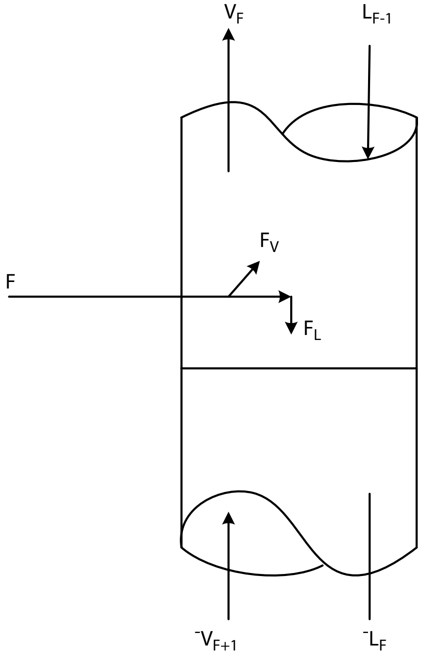


Figure 19.110 Condition at the feed.

12. Estimate the bubble-point temperature of the bottoms product B if not determined above.
13. The composition of $\bar{V}_{n+1}$ is the equilibrium composition with B and is calculated by $y_{i_{n+1}} = (K_i x_i)_{n+1}$.
14. Since the quantities of $\bar{L}_n$ and $\bar{V}_{n+1}$ are known, the composition of $\bar{L}_n$ can be determined by material balance:

$$x_{i_n} = \frac{\bar{V}_{n+1} y_{n+1} + Bx_{i_B}}{\bar{L}_n} \tag{19.485}$$

15. The temperature of plate n is the bubble-point temperature of L_n and is calculated in the same manner as the reboiler temperature.
16. The calculations are then continued, plate by plate, up the column; determining liquid compositions by material balance, temperature by bubble-point computation, and vapor composition by the equilibrium calculations until the liquid and vapor compositions approximate the compositions of the liquid and vapor potions of the feed, or until the key component ratio matches approximately that of the feed. This procedure and observations apply as to the computations proceeding down from the top of the column. The method is sufficient if the key component composition ratio approximately matches that of the feed and the plate temperature is essentially the same as that of the feed.

19.49 Multiple Feeds and Side Streams for a Binary Mixture

Simple binary distillation columns have been considered with single feed and two products, namely, the distillate and bottoms. However, it is possible to have more than one feed and more than two product streams. For example, in the distillation of crude petroleum, several products such as natural gasoline, liquefied petroleum gas (LPG) propane, butane, motor gasoline, solvent naphtha, aviation fuels, kerosene, diesel oil, gas oils, lube oils, fuel oils, and asphalts are all being withdrawn from the column at different points. The McCabe Thiele method can be used to analyze such complex configurations.

Two Feed Streams

Consider a typical distillation column as shown in Figure 19.111 having two feeds F_1 and F_2, with an overhead and bottoms product D and B, respectively. Figure 19.112 is the related McCabe–Thiele diagram showing the various operating lines. Each feed has its own q line, i.e., feed line. There are three distinct sections in Figure 19.111, namely:

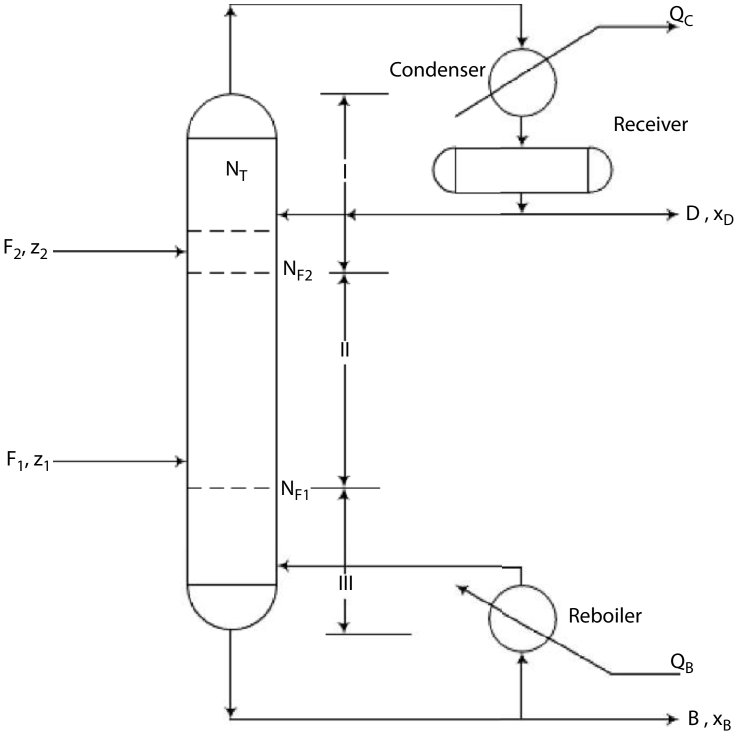


Figure 19.111 A two-feed column.

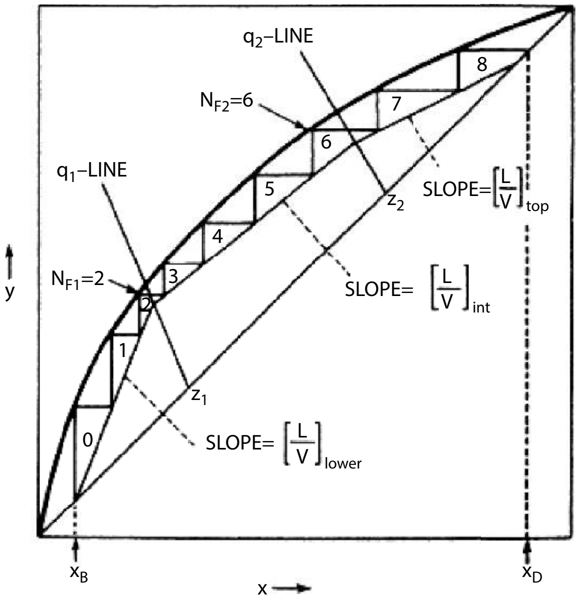


Figure 19.112 The McCabe–Thiele diagram of a two-feed column.

Section I. The top section above the top feed tray.
Section II. The intermediate section between the two feed trays.
Section III. The bottoms section below the lower feed tray.

There are three operating lines in Figure 19.112, each having different slope (i.e., L/V ratio).

Side Stream Columns

Some columns receive feed at one point but have one or more side streams drawing liquid or vapor product over the normal distillate and bottoms products—as in petroleum refineries. Figures 19.113a and d schematically represent a system in which a vapor or liquid side stream may be withdrawn. The removal of liquid from tray N_s reduces the liquid flow rate in the section below the side stream. Hence, the L/V ratio is higher above the side stream than below it as shown in Figure 19.113b. Three additional variables, such as flow rate, S, type of side draw (liquid or vapor), and location or composition x_S or y_S, must be specified. The operating equation for the middle section can be derived from the mass balances around the top or bottom section of the column. Further, the operating line changes the slope at the side stream tray. The material balance equations at the top and the side stream assuming constant molal overflow are

$$y=\frac{L}{V}x+\frac{D}{V}x_D \qquad (19.486)$$

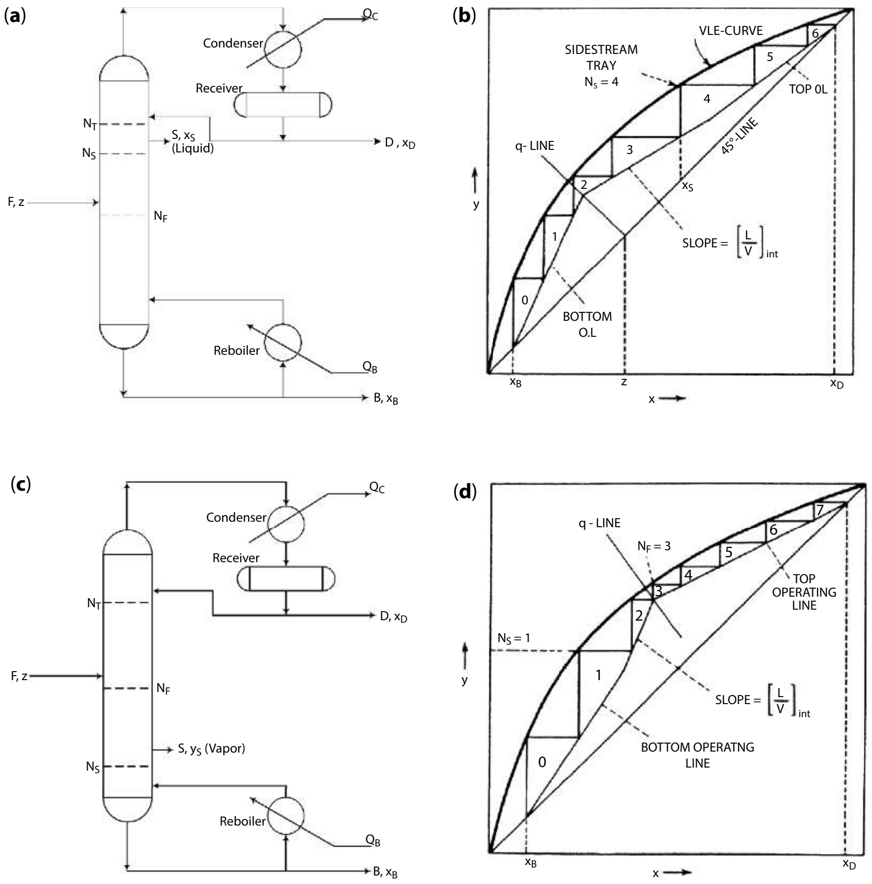


Figure 19.113 (a) A liquid side stream column. (b) The McCabe–Thiele diagram with a liquid side stream column. (c) A vapor side stream column. (d) The McCabe–Thiele diagram with a vapor side stream column.

and

$$y = \frac{L'}{V}x + \frac{Dx_D + Sx_S}{V} \tag{19.487}$$

Equating the two operating line Equations (19.486) and (19.487), the intersection occurs at

$$(L - L')x = Sx_S \tag{19.488}$$

and since

$$L = L' + S \tag{19.489}$$

the point of intersection becomes $x = x_S$.

The intersection of the lines is

$$y = \frac{L'}{V}x + \frac{Sx_S + Dx_D}{V} \tag{19.490}$$

and y = x intercept occurs at

$$x = \frac{Sx_S + Dx_D}{S + D} \tag{19.491}$$

Figure 19.113c schematically represents a system where a vapor side stream is withdrawn from trays N_S. The vapor flow rate in the section above N_S is lower than in the section below it. As the liquid flow rate L remains unchanged, the L/V ratio above the side stream is higher than below it (Figure 19.113d). For a vapor side stream, the feed line is horizontal ($q = 0$) at $y = y_S$. A balance on vapor flow rates gives

$$V' = V + S \tag{19.492}$$

As L remains unchanged, L'/V' can be calculated if L and V are known. If a specified value of x_S or y_S is desired, the problem becomes one of a trial and error.

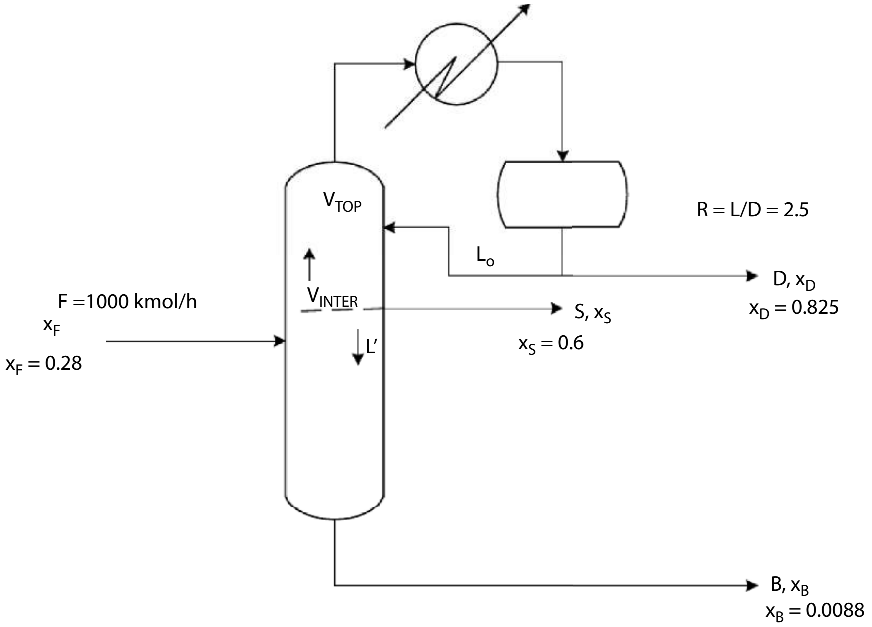
Example 19.45

An ethanol–water mixture is to be separated by continuous fractionation in a sieve-plate distillation column operated at 101.325 kPa. A feed flow rate of 1000 kmol/h containing 28 mol % ethanol is to be separated into two ethanol-rich streams: a distillate containing 82.5 mole % ethanol and a side stream containing 60 mol % ethanol. It is required to recover 98% of the ethanol in the feed in these two product streams.

The molal flow rate of the distillate will be twice that of the side stream. The feed will be liquid at its boiling point, and the product streams will be removed as liquid, also at their respective boiling points. An external reflux ratio of 2.5 to 1.0 will be applied at the top of the column. Stating clearly any assumptions, calculate the following:

- **a.** The composition of the bottoms product.
- **b.** Estimate the number of sieve plates required if the plate efficiency is 60%.
- **c.** Indicate briefly how a distillate containing 99.5 mole % ethanol could be obtained.

Solution



Assuming constant molal overflow:

a. Overall material balance:

$$F = D + S + B \tag{19.1}$$

The component balance on the more volatile component at the top of the column and side stream is

$$Fx_F = Dx_D + Sx_S \tag{19.2}$$

Since it is required to recover 98% of the ethanol in the feed, then

$$(0.98)\,(100)\,(0.28) = (0.825)\,D + 0.6S$$

Since the molar flow rate of the distillate is twice that of the side stream:

$$S = \frac{D}{2}$$

The feed is at its boiling point, q = 1 or q/(q-1) = ∞

$$274.4 = 0.825D + 0.3D$$

Hence,

$$D = 243.9 \text{ kmol/h}$$

and

$$S = 243.9/2 = 121.95 \text{ kmol/h}$$

Material balance on the upper operating line:

$$V = L + D$$

Component balance is

$$Vy = Lx + D\,x_D$$

The equation of the upper operating line is

$$y = \frac{L}{V}x + \frac{D}{V}x_D$$

where R = L/D or V = (R+1)D

$$y = \frac{R}{(R+1)}x + \frac{x_D}{(R+1)}$$

The intercept of the upper operating line (UOL) is

$$I = \frac{x_D}{(R+1)} = \frac{0.825}{3.5} = 0.236$$

The bottom flow rate is

$$\begin{aligned} B &= F - D - S \\ &= 1000 - 243.9 - 121.95 \\ &= 634.15 \text{ kmol/h} \end{aligned}$$

But 2% of the ethanol in the feed is removed at the bottoms, that is,

$$0.02\,Fx_F = B\,x_B$$

$$x_B = \frac{0.02 \times 1000 \times 0.28}{634.15} = 0.0088$$

b. The intermediate slope is L′/V′ assumed a constant molal overflow:

$$R = L_o/D = 2.5 \quad L_o = 2.5D = 2.5 \times 243.9 = 609.75 \text{ kmol/h}$$

Mass balance at the top of the column is

$$\begin{aligned} V_{TOP} &= L_o + D \\ &= 609.75 + 243.9 \\ &= 853.65 \text{ kmol/h} \end{aligned}$$

Side stream is liquid at its boiling point; therefore, a mass balance on the side stream is

$$V_{INTER} + L_o = S + L' + V_{TOP}$$

But $V_{INTER} = V_{TOP}$

Therefore,

$$\begin{aligned} L' &= L_o - S \\ &= 609.75 - 121.95 \\ &= 487.8 \text{ kmol/h} \end{aligned}$$

The slope of the side stream is

$$\frac{L'}{V'} = \frac{487.8}{853.65} = 0.571$$

The number of sieve plates from the McCabe–Thiele diagram (Figure 19.114) is 13. The actual number of sieve plates with a plate efficiency of 60% is

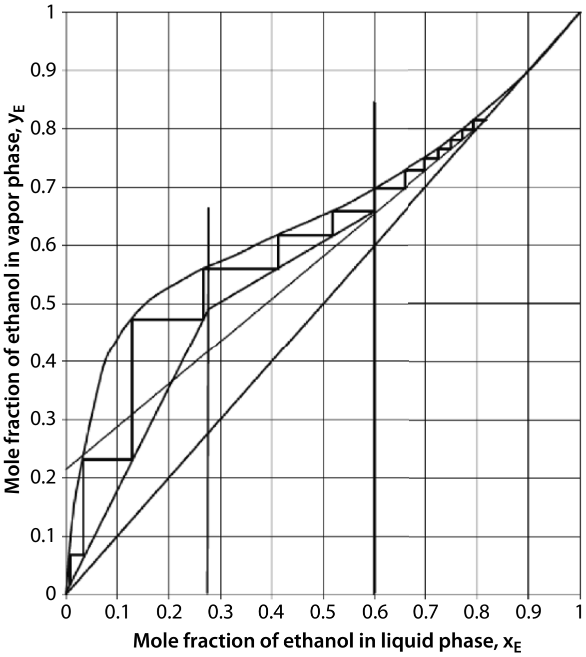


Figure 19.114 McCabe–Thiele diagram for Example 19.45.

$$\text{Actual number of plates} = \frac{\text{Number of plates}}{\text{Efficiency}}$$

$$= 13.0/0.60$$

$$= 21.6 \text{ sieve plates, i.e. } 22.$$

c. Use an azeotrope extracting agent, e.g., benzene, cyclohexane, or adjust the pressure to change the composition of the azeotrope. This will require a greater ΔP shift.

19.50 Chou and Yaws Method [183]

This method for multicomponent distillation involving more than one feed and more than one side stream requires a reliable minimum reflux.

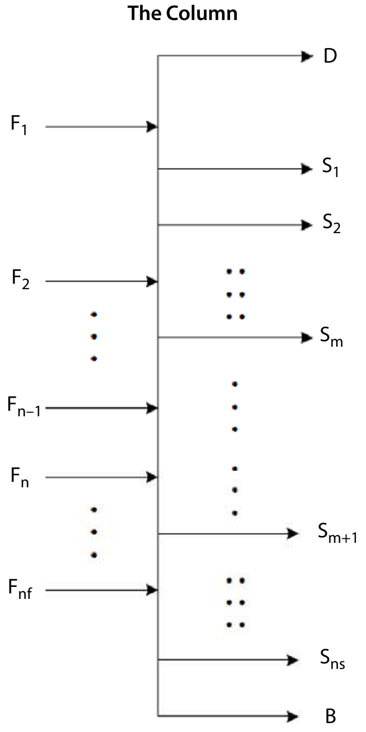
In summary, the calculation procedure is as presented by the authors.

For the systems rated above, the minimum reflux ratio is [183]

$$R_{min} = R_F + R_{OF} + R_s \tag{19.493}$$

$$R_{min} = R_F + \sum F_{FR,j} F_j + \sum F_{SR,k} S_k \tag{19.494}$$

This includes recognizing the contribution from the feed (R_F), "other feeds" (R_{OF}), and side streams (R_S). The R_F portion is determined assuming no other feeds or side streams are present. The R_{OF} and R_S parts represent the summation of the contributions of other feeds and side streams to the overall column minimum reflux ratio. The calculation sequence consists basically of three steps, here reproduced by permission of Chemical Engineering, Chou and Yaws, April 25, 1988, all rights reserved [183]:



1. Determine Underwood θ for each feed using equation involving feed concentration:

$$\theta_1, \theta_2, \ldots\ldots\ldots\ldots\theta_{nf}$$

$$1 - q_{Fn} = \sum_{i=1}^{c} \frac{\alpha_i Z_i, F_n}{(\alpha_i - \theta_n)} \tag{19.475}$$

2. Determine the "minimum reflux ratio" candidate for each feed:

$$R_{min,1}, R_{min,2}, \ldots\ldots\ldots\ldots\ldots, R_{min,nf}$$

$$\begin{aligned} R_{min,n} &= R_F + R_{OF} + R_S \\ &= R_{F,n} + \sum_{j=1}^{n-1} F_{FR,j} F_j + \sum_{k=1}^{m} F_{SR,k} S_k \end{aligned} \tag{19.496}$$

where:

$$R_{F,n} = \sum_{i=1}^{c} \frac{\alpha_i x_{i,D}}{\alpha_i - \theta_n} - 1$$

$$F_{FR,j} = -\frac{1}{D}\left(\sum_{i=1}^{c} \frac{\alpha_i Z_i F_j}{\alpha_i - \theta_n} + q_{F_j} - 1 \right)$$

$$F_{SR,k} = \frac{1}{D}\left(\sum_{i=1}^{c} \frac{\alpha_i Z_i S_k}{\alpha_i - \theta_n} + q_{S_k} - 1 \right)$$

3. Compare the candidates for minimum reflux ratio. The candidate having the largest (maximum) numerical value is the minimum reflux ratio for the column.

$$R_{min} = \text{Max}(R_{min,1}, R_{min,2}, \ldots\ldots\ldots\ldots\ldots, R_{min,nf})$$

Example 19.46: Distillation with Two Side Streams Feeds

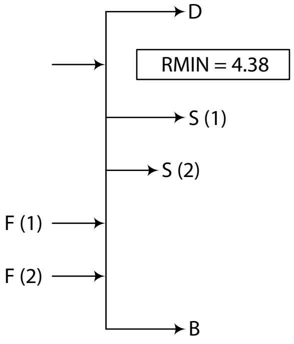
Data for Example 19.46, which includes two side streams

		Mole Fraction				
		Feed		**Distillate**	**Side stream**	
Component	**Relative Volatility α**	**z_i, F_1**	**z_i, F_2**	**x_D**	**z_i, S_1**	**z_i, S_2**
1 (LK)	2.0	0.38	0.45	0.999	0.975	0.900
2 (HK)	1.0	0.20	0.30	0.001	0.025	0.099
3 (HK + 1)	0.5	0.22	0.25	0	0	0.001
Feeds: F_1 = 50 mol/h, q_{F1} = 1 (saturated liquid) F_2 = 100 mol/h, q_{F2} = 0 (saturated vapor) Side streams: S_1 = 20 mol/h, q_{S1} = 1 (saturated liquid) S_2 = 20 mol/h, q_{S2} = 1 (saturated liquid) Distillate: D = 36 mol/h						

Minimum reflux and other results for Example 19.46

1.	UNDERWOOD THETAS:	
	FOR FEED 1	THETA (1) = 1.164
	FOR FEED 2	THETA (2) = 1.485
2.	MINIMUM REFLUX CANDIDATES:	
	FOR FEED 1	RMIN (1) = 3.450271
	FOR FEED 2	RMIN (2) = 4.375502
3.	TRUE MINIMUM REFLUX RATIO:	
	RMIN = 4.38	

Column representation of results of Example 19.46



where

B = bottoms flow rate, mol/h
c = number of components
D = distillate flow rate, mol/h
F = flow rate of feed, mol/h
F_j = flow rate of feed j, mol/h
F_{FR} = factor for contribution of other feed flow to minimum reflux
F_{FRj} = factor for contribution of feed j flow to minimum reflux
F_{SR} = factor for contribution of side stream flow to minimum reflux
$F_{SR,k}$ = factor for contribution of side stream k flow to minimum reflux
HK = heavy key component
L = liquid flow rate, mol/h
LK = light key component
nf = number of feeds
ns = number of side streams
m = number of side streams above feed n
q_F = thermal condition of feed
q_S = thermal condition of side stream
R = reflux ratio
R_F = feed component of minimum reflux
$R_{F,n}$ = feed component of minimum reflux for feed n
R_{OF} = other feeds component of minimum reflux
R_{min} = minimum reflux ratio
R_S = side stream component of minimum reflux
S = flow rate of side stream: n, mol/h
S_k = flow rate of side stream k, mol/h
V = vapor flow rate, mol/h
x_i = mole fraction of component i in liquid
y_i = mole fraction of component i in vapor
$z_{i,F}$ = mole fraction of component i in feed
$z_{i,Fj}$ = mole fraction of component i in feed j
$z_{i,S}$ = mole fraction of component i in side stream
$z_{i,Sk}$ = mole fraction of component i in side stream k
α = relative volatility
θ = underwood parameter

Subscripts

B = bottoms
D = distillate
F = feed
F_j = feed j
F_n = intermediate feed

19.51 Optimum Reflux Ratio and Optimum Number of Trays Calculations

van Winkle and Todd have developed shortcut graphical methods that are very useful for estimating the optimum reflux ratio R_{Ropt} and optimum number of stages N_{opt}, such as to minimize operating and investment costs for a given separation of multicomponent systems in a fractionation operation. Their methods were based on the following:

1. Using the shortcut methods of Fenske for the minimum number of stages as shown in Figure 19.115.

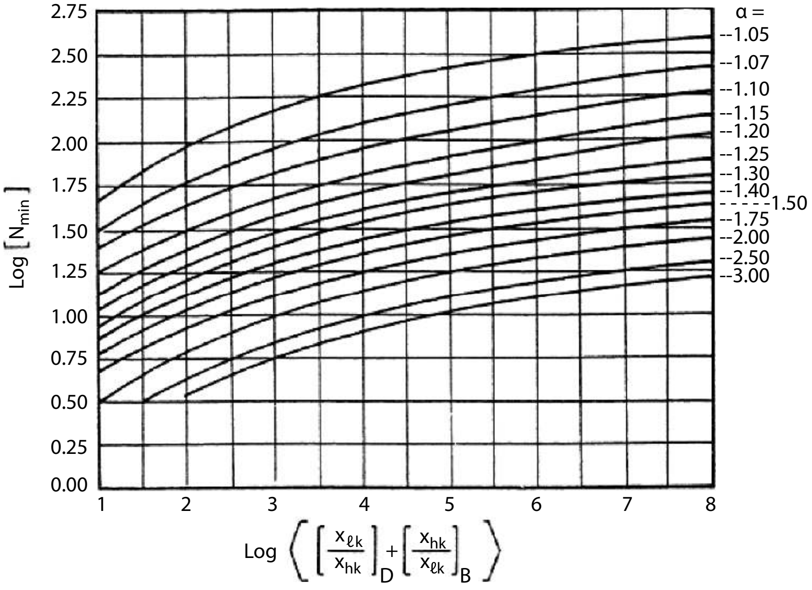


Figure 19.115 Graphical representation of Fenske equation for estimating minimum plates (Source: P. Chattopadhyay [184]).

2. Using the Underwood method for the minimum reflux ratio (Figures 19.116, 19.117, 19.118, and 19.119).
3. Using Gilliland's correlation for determining the relationship between theoretical stages and reflux ratio to their minimum value.

This method has been tested on several hundred cases of multicomponent distillation in which $\alpha,(x_{LK}/x_{HK})_F$, percent recovery of LK in the distillate, and percent recovery of the HK in the bottoms were varied over a practical range. The following observations were made about the result:

Effect of Pressure: Pressure exerts no appreciable effect on the optimum values of total number of trays and reflux ratio.

Effect of Feed Rate: The optimum reflux ratio RR_{opt} and optimum number of stages N_{opt} vary with the changes in feed flow rates. The value of $(N/N_{min})_{opt}$ known as the tray factor increases with the increase in feed rate, while the value of $(RR/RR_{min})_{opt}$ known as the reflux factor decreases at the same time.

Material of Construction: The materials of construction mainly used are plain carbon steel and stainless steel; Monel metal of tower and trays has a direct effect on $(N/N_{min})_{opt}$ and $(RR/RR_{min})_{opt}$. However, with the increase in the cost of material M_c, the optimum values of (N/N_{min}) and (RR/RR_{min}) decrease.

Utility Cost: The inclusion of utility cost (e.g., cost of steam + cost of cooling water + cost of electricity) in reducing the total cost greatly affects the RR_{opt} and N_{opt}. When the utility cost (also known as the operating cost) is taken into account, the optimum level of (N/N_{min}) always increases, whereas the optimum level of (RR/RR_{min}) always decreases.

Correlations

The overall procedure is based on the correlation of $(N/Nmin)_{opt}$ and $(RR/RR_{min})_{opt}$ for a given reference case. A base case is selected and then correction factors are correlated to take into account the effects of feed rates, cost of material, and the number of years of equipment amortization.

Originally, the selected base case is:

A 3-m-diameter column of plain carbon steel as the material of construction, and operating pressure of 696 kPa and 2-year amortization. The optimum value of (N/N_{min}) for both the reference case and the range of conditions was

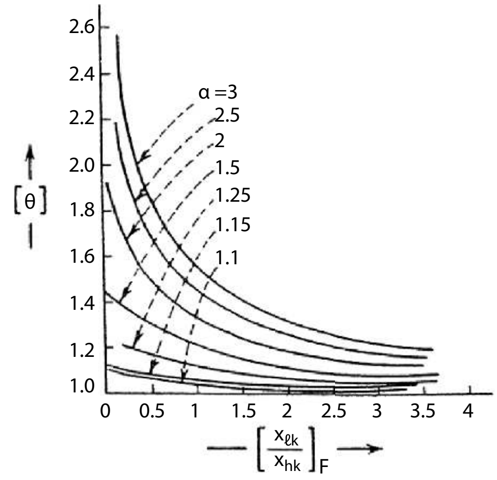


Figure 19.116 Underwood's parameter (θ) as a function of the ratio of the keys in the feed (Source: P. Chattopadhyay [184]).

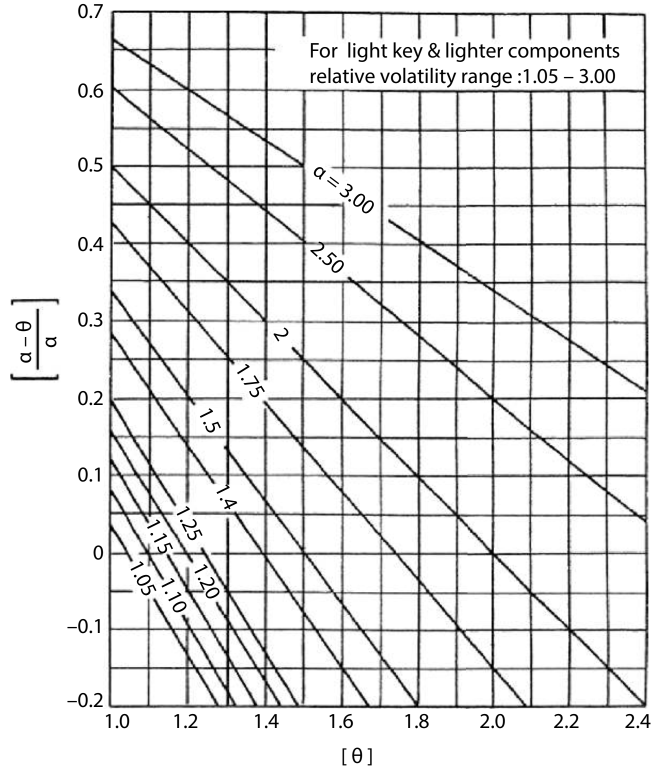


Figure 19.117 Dependence of $\left[\frac{\alpha-\theta}{\alpha}\right]$ on θ for the light key and lighter components for calculating Underwood's (RR_m + 1). Relative volatility α, 1.05 < α < 3.0 (Source: P. Chattopadhyay [184]).

correlated as a function of the Fenske separation factor, using α as the parameter as illustrated in Figure 19.120 to estimate $(N/N_{min})_{ref}$ values.

To account for the effects of various feed rates, the feed correction factor ϕ was correlated as a function of feed rate, using α as the parameter as shown in Figure 19.121. Figure 19.121 is used to evaluate the correction factor ϕ for $(N/N_{min})_{ref}$ and ϕ is used to calculate $(N/N_{min})_{opt}$ as

$$\left(\frac{N}{N_{min}}\right)_{opt} = \phi \cdot \Gamma \cdot \left(\frac{N}{N_{min}}\right)_{ref} \tag{19.497}$$

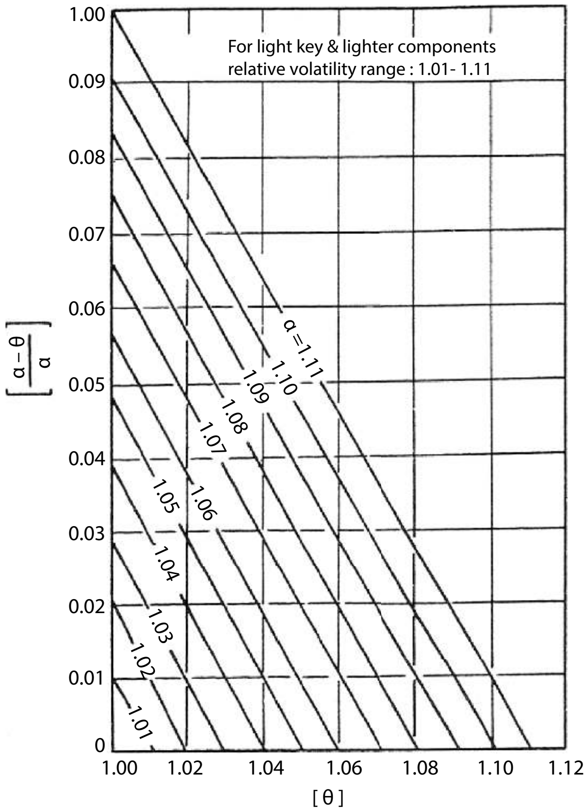


Figure 19.118 Dependence of $\left[\frac{\alpha-\theta}{\alpha}\right]$ on θ for the light key and lighter components for calculating Underwood's (RR_m + 1). Relative volatility α, $1.01 < \alpha < 1.11$ (Source: P. Chattopadhyay [184]).

Γ is the correction factor for $(N/N_{min})_{ref}$ for the depreciation period and material of construction. This correction factor is obtained from Figure 19.122, which plots Γ as a function of (n/M_c) for various values of α. $(RR/RR_{min})_{opt}$ is obtained from Figure 19.123, and $(N/N_{min})_{ref}$ correction factor for feed ϕ is obtained from Figure 19.124; $(RR/RR_{min})_{opt}$ correction factor for amortization Δ is determined from Figure 19.125.

Procedure

In order to determine the optimum combination of reflux ratio and the number of stages based on minimum yearly operating cost, the following parameters are required:

1. Feed rate.
2. Feed condition and feed properties.
3. Operating pressure.
4. Operating temperatures.
5. Relative volatilities of the components.
6. Specified separation of key components.
7. Material costs.
8. Amortization years.

The algorithm is as follows:

1. Determine the value of N_{min}. Use either Figure 19.115 or Fenske equation (19.182).

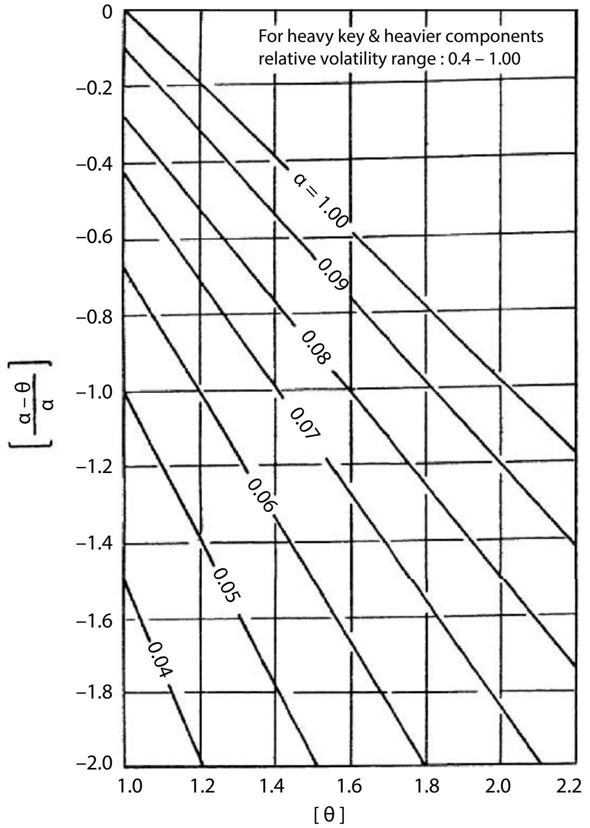


Figure 19.119 Dependence of $\left[\frac{\alpha-\theta}{\alpha}\right]$ on θ for the light key and lighter components for calculating Underwood's (RR_m + 1). Relative volatility α, 0.4 < α < 1.0 (Source: P. Chattopadhyay [184]).

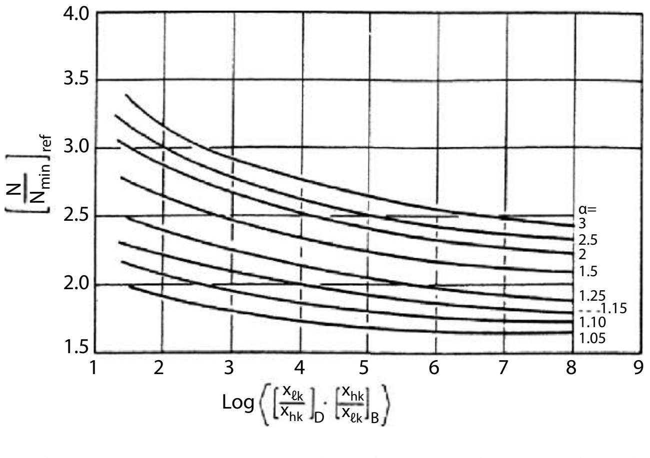


Figure 19.120 Correlation of the ratio of optimum to minimum number of trays as a function of Fenske separation factor (Source: P. Chattopadhyay [184]).

2. Determine RRmin from Underwood's method. Use Figure 19.116 to determine the Underwood's parameter θ. Alternatively use the spreadsheet program Example 19.41.xlsx. Use Equations (19.373) and (19.374) or Figures 19.117, 19.118, and 19.119 to determine RR_{min}.
3. Obtain the reference value $(N/N_{min})_{ref}$ from Figure 19.120 using Fenske separation factor for the correct α.

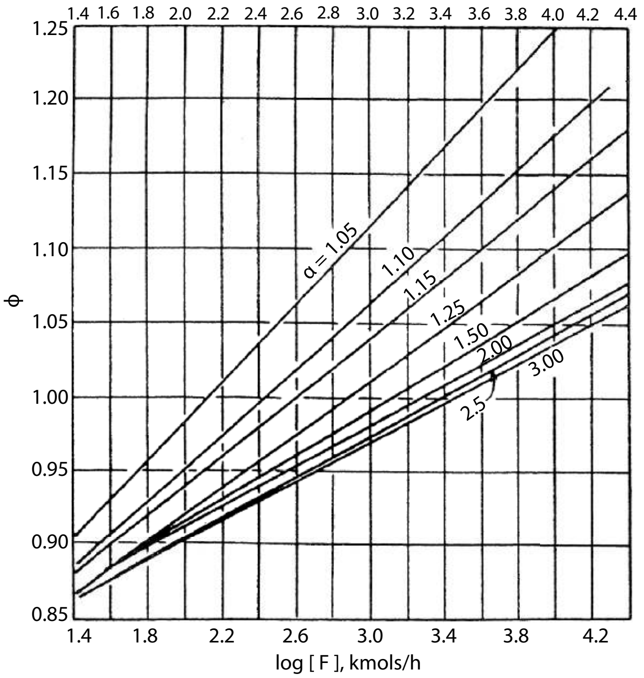


Figure 19.121 Feed rate correction factor for $(N/N_{min})_{ref}$ (Source: P. Chattopadhyay [184]).

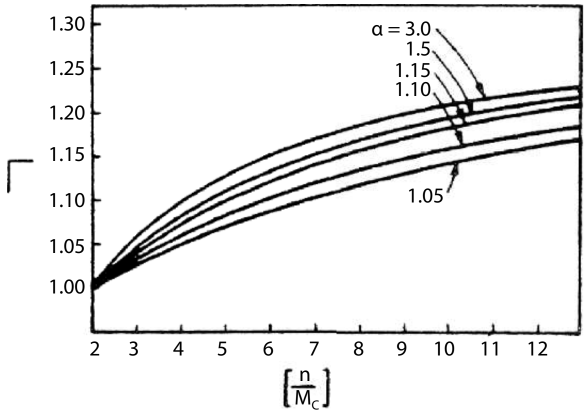


Figure 19.122 Correction factor for $(N/N_{min})_{ref}$ for the number of years amortized (n), divided by the cost of materials (Mc) (Source: P. Chattopadhyay [184]).

4. Determine the feed correction factor ϕ from Figure 19.121. Use the correct feed rate (F) and α.
5. The amortization time (n) and the material cost factor (M_c) are used to obtain the (N/N_{min}) correction factor (Γ) for depreciation and material of construction, from Figure 19.122. Calculate the optimum value of N/N_{min} from Equation (19.497):

$$\left(\frac{N}{N_{min}}\right)_{opt} = \phi \cdot \Gamma \cdot \left(\frac{N}{N_{min}}\right)_{ref}$$

6. Using the value of N_{min} from step 1 and the value of the RHS of Equation (19.497) in step 5, determine the optimum value of the number of theoretical stages required.

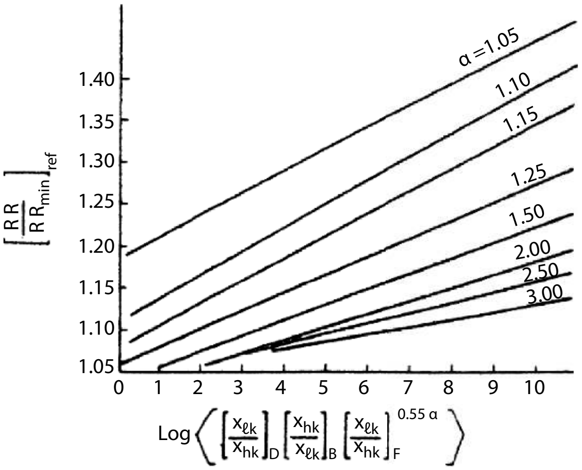


Figure 19.123 Correlation of optimum is to minimum reflux ratio for the reference case $(RR/RR_{min})_{ref}$ (Source: P. Chattopadhyay [184]).

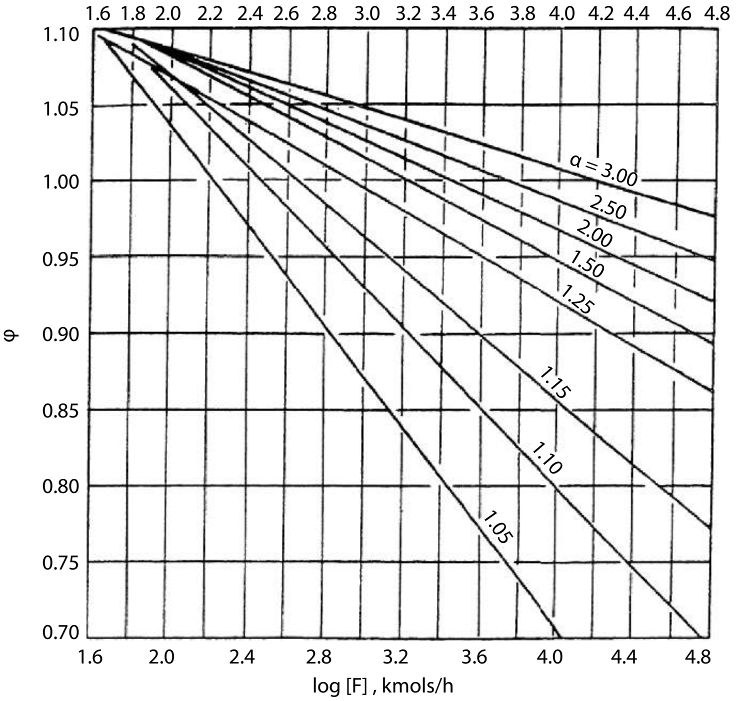


Figure 19.124 Feed rate correction factor for $(RR/RR_{min})_{ref}$ (Source: P. Chattopadhyay [184]).

7. From Equation (19.123), determine $(RR/RR_{min})_{ref}$ on the basis of empirical separation factor (plotted in the abscissa).
8. From Figure 19.124, determine (RR/RR_{min}) correction factor φ for the feed. Use the correct feed rate (F) and α.
9. For the selected amortization time (n) and material cost factor (M_c), read out the correction factor Δ for (RR/RR_{min}) from Figure 19.125. Calculate the optimum value of (RR/RR_{min}) from

$$\left(\frac{RR}{RR_{min}}\right)_{opt} = \phi \cdot \Delta \cdot \left(\frac{RR}{RR_{min}}\right)_{ref} \tag{19.498}$$

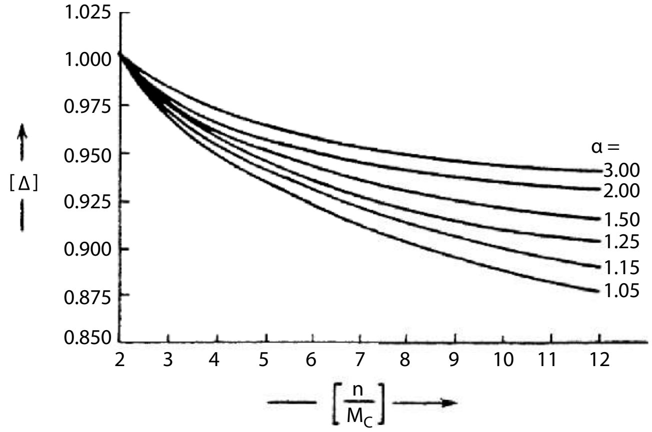


Figure 19.125 Correction factor for $(RR/RR_{min})_{ref}$ for the depreciation period and material of construction (Source: P. Chattopadhyay [184]).

Table 19.45 Economic R/R_{min} and N/N_{min} vs. heat sink source.

Method of condensing the reflux	R/R_{min}	N/N_{min}
Low-level refrigeration (-300 to 150°F)	1.05 – 1.10	2.0 – 3.0
High-level refrigeration (-150 to 50 °F)	1.10 – 1.20	1.8 – 2.0
Cooling water at cost of circulation and limited treating	1.20 – 1.5	1.0 – 1.8
Air cooling	1.40 – 1.5	1.6 – 1.8

10. With RR_{min} from step 2 and the value of the RHS of Equation (19.498), the optimum value of the reflux ratio RR_{opt} is determined.

Generally, the optimum reflux ratio for a multicomponent system is obtained by comparing operating (energy) costs with the capital costs and then optimizing the two. Results of previous economic designs have been compiled, compared, and reduced to a relationship of (RR/RR_{min}) or equivalent N/N_{min} based on the type of heat sink employed. Table 19.45 shows the economic ranges of reflux to minimum reflux and the resulting values of stages to minimum stages.

Example 19.47: Minimum and Optimum Values of Theoretical Trays and Reflux Ratio

A multicomponent system comprising of four hydrocarbons: isobutane (iC_4H_{10}), n-butane (nC_4H_{10}), isopentane (iC_5H_{12}), and n-pentane (nC_5H_{12}), is to be separated as specified in the table.

Component	Feed moles	Equilibrium constant K	Relative volatility α	Distillate moles	Bottoms moles
iC_4H_{10}	0.06	2.15	2.5749	0.0596	0.0004
nC_4H_{10} (LK)	0.17	1.70	2.0359	0.1615	0.0085
iC_5H_{12} (HK)	0.32	0.835	1.0000	0.0160	0.3040
nC_5H_{12}	0.45	0.70	0.8383	0.0054	0.4446
Total	1.00			0.2425	0.7575

The column operating pressure = 103 kPa
Top temperature = 420 K
Bottom temperature = 470K
95% recovery of both keys is desired.
Feed is saturated liquid at its boiling point (q = 1)
Take: n-butane (nC_4H_{10}) as light key component
i-pentane (iC_5H_{12}) as heavy key component.

Input Data

Material of construction	= Carbon steel for all equipment
Material cost factor	= 1
Depreciated period	= 10 years
Feed rate	= 500 kmol/h
Tray efficiency	= 85%

Determine the following: N_{min}, N_{opt}, R_{min}, and R_{opt}.

Solution

Table 19.46 gives the Excel spreadsheet (Example 19.47.xlsx) results of Example 19.47.

From Table 19.44:

Minimum reflux ratio, RR_{min}	= 2.8036
Minimum number of stages, N_{min}	= 8.3
Underwood constant, θ	= 1.6213
Relative volatility, α_{LK}/α_{HK}	= 2.0359

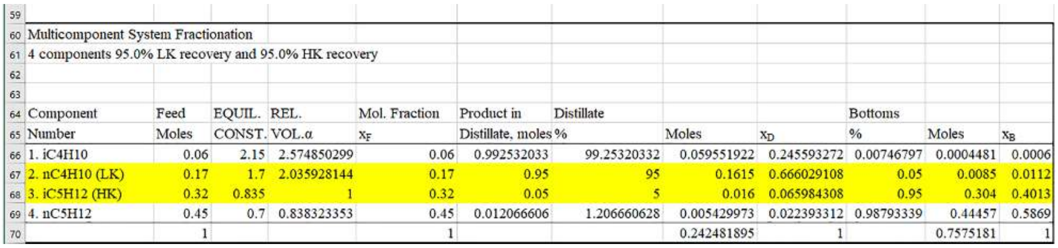
The following procedure is used to calculate N_{opt} and R_{opt}:

1. Graphical representation of Fenske equation for estimating minimum plates as:

$$\log\left[\left(\frac{x_{LK}}{x_{HK}}\right)_D\left(\frac{x_{HK}}{x_{LK}}\right)_B\right] \text{ from Figure 19.120:}$$

$$\log\left[\left(\frac{0.6660}{0.066}\right)_D\left(\frac{0.4013}{0.0112}\right)_B\right] = 2.558$$

Table 19.46 Results for multicomponent system fractionating of Example 19.47.



Multicomponent System Fractionation
4 components 95.0% LK recovery and 95.0% HK recovery

Component	Feed	EQUIL.	REL.	Mol. Fraction	Product in	Distillate			Bottoms		
Number	Moles	CONST.	VOL.α	x_F	Distillate, moles	%	Moles	x_D	%	Moles	x_B
1. iC4H10	0.06	2.15	2.574850299	0.06	0.992532033	99.25320332	0.059551922	0.245593272	0.00746797	0.0004481	0.0006
2. nC4H10 (LK)	0.17	1.7	2.035928144	0.17	0.95	95	0.1615	0.666029108	0.05	0.0085	0.0112
3. iC5H12 (HK)	0.32	0.835	1	0.32	0.05	5	0.016	0.065984308	0.95	0.304	0.4013
4. nC5H12	0.45	0.7	0.838323353	0.45	0.012066606	1.206660628	0.005429973	0.022393312	0.98793339	0.44457	0.5869
	1			1			0.242481895	1		0.7575181	1

The feed condition, Q	1.0
The heavy key component number	3
Percentage recovery of the light key component in the distillate (%)	95
Percentage recovery of the heavy key component in the bottoms (%)	95
Total moles in the distillate, moles	0.2425
Total moles in the bottoms, moles	0.7575
Total moles in the feed, moles	1.0
Underwood constant (θ)	1.6214
Factor for the reflux ratio	1.3
Minimum reflux ratio	2.804
Actual reflux ratio	3.65
Minimum number of stages	8.3
Number of theoretical plates in the column	16.2
The position of the feed plate	6.7

2. Underwood constant, θ = 1.6213

$$\left(\frac{x_{LK}}{x_{HK}}\right)_F = \frac{0.17}{0.32} = 0.5313$$

3. Correlation of the ratio of optimum to minimum number of trays as a function of Fenske separation factor from Figure 19.120.

$$\text{At } \log\left[\left(\frac{0.6660}{0.066}\right)\left(\frac{0.4013}{0.0112}\right)\right] = 2.558 \text{ and}$$

$$\alpha_{LK}/\alpha_{HK} = 2.0359$$

$$\left(\frac{N}{N_{min}}\right)_{ref} = 2.75$$

4. Correction factor ϕ for feed rate using Figure 19.121.

$$F = 500 \text{ kmol/h}$$

$$\log(F) = \log(500) = 2.698$$

$$\text{From Figure 19.70, } \phi = 0.97$$

5. Correction factor Γ for $(N/N_{min})_{ref}$ for the number of years amortized (n), divided by the cost of materials (M_c).

$$n = 10 \text{ years, } M_c = 1.0$$

$$\frac{n}{M_c} = \frac{10}{1} = 10, \alpha_{LK}/\alpha_{HK} = 2.0359$$

$$\text{Figure 19.122, } \Gamma = 1.185$$

6. Optimum value of $\frac{N}{N_{min}}$ is

$$\begin{aligned}\left(\frac{N}{N_{min}}\right)_{opt} &= \phi \bullet \Gamma \bullet \left(\frac{N}{N_{min}}\right)_{ref}\\ &= (0.97)(1.185)(2.75)\\ &= 3.16\end{aligned}$$

7. Optimum number of theoretical stages, where $N_{min} = 8.3$ (Table 19.47).

$$\begin{aligned}N_{opt} &= N_{min}\left(\frac{N}{N_{min}}\right)_{opt}\\ &= (8.3)(3.16)\\ &= 26.24\end{aligned}$$

8. Correlation of optimum to minimum reflux ratio for the reference case $(RR/RR_{min})_{ref}$ using composition values of LK and HK in Table 19.44.

$$\begin{aligned}&\log\left[\left(\frac{x_{LK}}{x_{HK}}\right)_D \left(\frac{x_{HK}}{x_{LK}}\right)_B \left(\frac{x_{LK}}{x_{HK}}\right)_F^{0.55\alpha}\right]\\ &= \log\left[\left(\frac{0.666}{0.066}\right)_D \left(\frac{0.4013}{0.0112}\right)_B \left(\frac{0.17}{0.32}\right)^{(0.55)(2.0359)}\right]\\ &= 2.25058\end{aligned}$$

From Figure 19.123, $\alpha_{LK}/\alpha_{HK} = 2.0359$

Table 19.47 Economic optimum reflux ratio for typical petroleum fraction distillation near 1 atm[a].

	Factor for optimum reflux										
	$f = (R_{opt}/R_m) - 1$				**Factor for optimum trays**						
	$R_{opt} = (1 + f) R_m$				**N_{opt} / N_m**						
		$N_m = 10$			**$N_m = 20$**		**$N_m = 50$**		**$N_m = 10$**	**$N_m = 20$**	**$N_m = 50$**
		Rm			**R_m**		**R_m**		**R_m**	**R_m**	**R_m**
	1	**3**	**10**	**1**	**3**	**10**	**1**	**10**	**1 to 10**	**1 to 10**	**1 to 10**
Base case	0.20	0.12	0.10	0.24	0.17	0.16	0.31	0.21	2.4	2.3	2.1
Payout time 1 yr.	0.24	0.14	0.12	0.28	0.20	0.17	0.37	0.24	2.2	2.1	2.0
Payout time 5 yr.	0.13	0.09	0.07	0.17	0.13	0.10	0.22	0.15	2.7	2.5	2.2
Steam cost \$0.30/M lb	0.22	0.13	0.11	0.27	0.16	0.14	0.35	0.22	2.3	2.1	2.0
Steam cost \$0.75 /M lb	0.18	0.11	0.09	0.21	0.13	0.11	0.29	0.19	2.5	2.3	2.1
G_a = 50 lb mole/(hr)(sq.ft)	0.06	0.04	0.03	0.08	0.06	0.05	0.13	0.08	3.1	2.8	2.4

[a]The "base case" is for payout time of 2 yr, steam cost of \$0.05/1000 lb, vapor flow rate Ga = 15 lb mol/(hr)(sq. ft.). Although the capital and utility costs are prior to 1975 and are individually far out of date, the relative costs are roughly the same so the conclusions of this analysis are not far out of line. Conclusion: For systems with nearly ideal VLE, R is approx. $1.2R_{min}$ and N is approx. $2.0\ N_{min}$.
Source: Happen and Jordan, Chemical Process Economics, Decker, New York, 1975.

$$\left[\frac{RR}{RR_{min}}\right]_{ref} = 1.06$$

9. Correction factor φ for feed rate from Figure 19.124,

$$\log(F) = 2.70 \text{ at } \alpha_{LK}/\alpha_{HK} = 2.0359$$

$$\varphi = 1.04$$

10. Depreciation and material of construction correction factor.

$$n = 10 \text{ years}, M_c = 1.0, n/M_c = 10, \ \alpha_{LK}/\alpha_{HK} = 2.0359$$

$$\text{From Figure 19.74}, \Delta = 0.9375$$

$$\left(\frac{RR}{RR_{min}}\right)_{opt} = \phi \cdot \Delta \cdot \left(\frac{RR}{RR_{min}}\right)_{ref}$$
$$= 1.0335$$

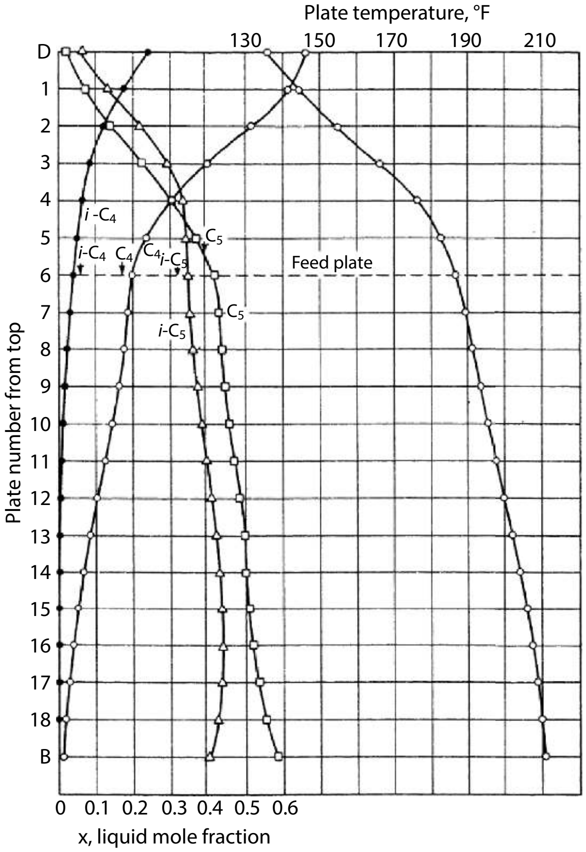


Figure 19.126 Typical concentration profiles in separation of light hydrocarbon mixtures when no substantial inversions of relative volatilities occur (Source: van Winkle, Distillation, McGraw-Hill, New York, 1967).

11. Optimum reflux ratio, RR_{opt}, where RR_{min} = 2.8036 (Table 19.46)

$$
\begin{aligned}
RR_{opt} &= RR_{min}\left(\frac{RR}{RR_{min}}\right)_{ref} \\
&= (2.8036)(1.0335) \\
&= 2.897
\end{aligned}
$$

Generally, the reflux ratio affects the cost of the column, both in the number of trays and the diameter. It also affects the cost of operation, i.e., the costs of heat and cooling supply and power for the reflux pump. Therefore, the basis for choice of an optimum reflux ratio involves an economic balance. Table 19.47 summarizes an economic balance of optimum reflux ratio for typical petroleum fractional distillation. The conclusion may be drawn that the optimum reflux ratio is about 1.2 times the minimum and also the number of trays is about twice the minimum. Figure 19.126 shows the temperature profiles of iC_4, nC_4, iC_5, and nC_5 through the column using Sorel method of plate-to-plate calculation.

Flow Regime in Trays

Flow regimes in a trayed distillation column are based on their vapor/liquid rates. The most encountered are as follows:

1. Froth regime: normal liquid rate.
2. Spray regime: low liquid rate.
3. Emulsion regime: encountered in high-pressure systems with relatively high liquid loads. Most of the gas is emulsified within the liquid, and the mixture behaves as a uniform two-phase fluid.

19.52 Tower Sizing for Valve Trays

Many types of trays are used in both fractionating and absorption columns. In a fractionating column, the bubble caps with the weirs and downcomers maintain a liquid level on the trays. The liquid flows across the tray via the downcomer and across the next tray in the opposite direction, while vapor flows up through the cap and the slots, thus mixing with the liquid. Figures 19.127 and 19.128 show a section of the column distributor and different types of packings. Figure 19.129 illustrates the vapor flows through the bubble cap, sieve, and valve trays. The riser in the bubble cap can help to prevent liquid from "weeping" through the vapor passage. Sieve or valve trays control weeping by vapor velocity. The bubble cap tray has the highest turn down ratio with the design of 8:1 to 10:1 ratio. Such trays are commonly used in glycol dehydrating columns [185].

Valve and sieve trays are favored because of lower cost and increased capacity compared to bubble cap trays for a given tower diameter. Sieve or perforated trays are plates with holes for vapor passage. Their simple construction makes them the least expensive of the three designs. Generally, the sieve tray has a higher capacity, but at low vapor rates, they are susceptible to "weeping" or dumping of the liquid through the holes. Furthermore, sieve trays are very good in fouling application or when solids are present. They can have large holes because they are highly resistant to clogging and are easy to clean. In addition, its turndown ratio is limited.

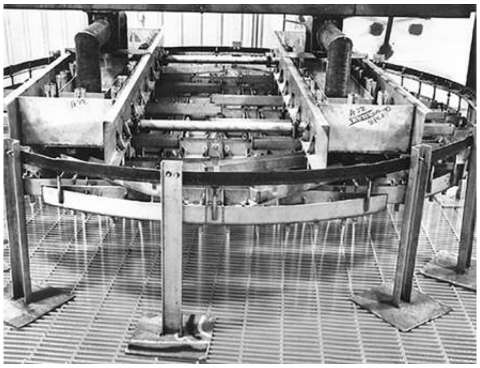


Figure 19.127 Nutter Engineering MTS-109 multi-pan two-stage liquid distributor. Courtesy of Nutter Engineering.

Figure 19.128 Nutter ring random packing. Courtesy of Nutter Engineering.

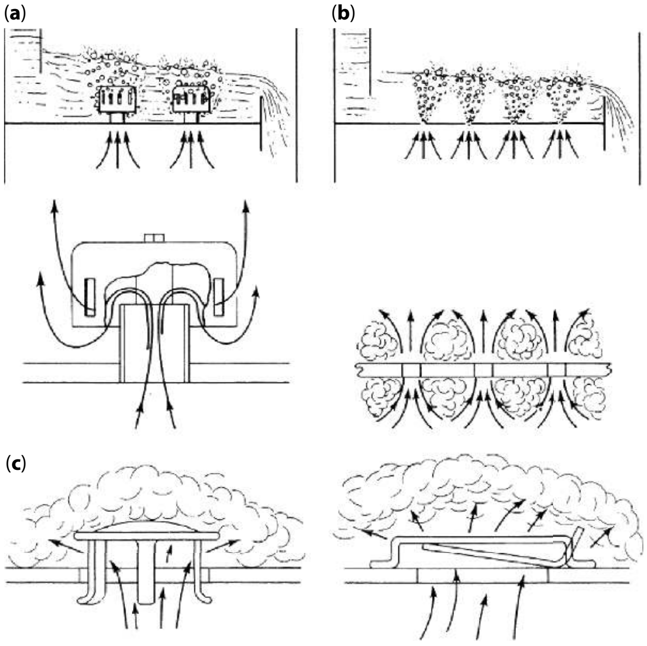


Figure 19.129 Flow through vapor passages. (a) Vapor flow through bubble cap. (b) Vapor flow through perforations. (c) Vapor flow through valves (Source: van Winkle, Distillation, McGraw-Hill, New York, 1967).

Trayed columns give satisfactory operation over a wide range of liquid and vapor loadings. Correspondingly, valve trays can maintain high efficiencies over a wider range of operating liquid and gas rates than with sieve trays. Valve units are more mechanically complex than sieve trays and therefore are more expensive to make. The liquid and vapor rates can vary independently over a broad range in the column resulting in a satisfactory operation. At low vapor rates, unsatisfactory tray dynamics may be evidenced by dumping of liquid, uneven distribution, and vapor pulsation. Alternatively, at high vapor rates, the tower floods as the liquid backs up in the downcomers. Poor vapor–liquid contact can result at low liquid rates. Correspondingly, at high liquid rates, flooding and dumping can also result as the liquid capacities are exceeded in the downcomers. Valve trays are designed to have better turndown properties than sieve trays and therefore are more flexible for varying feed rate. At the design vapor rate, valve trays have about the same efficiency as sieve trays. They can be designed for a lower pressure drop (ΔP) than sieve trays. However, they are susceptible to fouling or plugging if dirty solutions are distilled.

In these examples, valve trays are used for sizing the columns. The procedure uses the Tower sizing graphs and tables from Nutter [186] and Blackwell's [187] correlation.

The Equations

The following expressions are used for the design of valve trays columns:

The density radical R_D is defined by

$$R_D = \left(\frac{\rho_V}{\rho_V - \rho_L}\right)^{0.5} \qquad (19.499)$$

The velocity at zero liquid load is

$$V = \exp[A + B(\ln X) - C(\ln X)^2 + D(\ln X)^3 - E(\ln X)^4] \qquad (19.500)$$

where

$$X = \frac{\sigma}{\rho_V} \tag{19.501}$$

A, B, C, and D are constants
A = 0.22982
B = 0.44605
C = 0.03452
D = 0.00415
E = 0.00017

Equation (19.500) is valid for $0.1 \le X \le 3000$.

The tray spacing factor T_{SF} is defined by

$$T_{SF} = G + H(\ln X) \tag{19.502}$$

where Equation (19.502) is valid for $0.1 \le X \le 5000$. The constants G and H for different tray spacing are given in Table 19.48. Figure 19.130 illustrates the plots of T_{SF} versus $X(\sigma/\rho_V)$ for different tray spacing.

The operating volumetric vapor flow rate Q_V is

$$Q_V = \frac{W_V}{(3600)(\rho_V)}, \text{ft}^3/\text{s}. \tag{19.503}$$

The operating volumetric liquid flow rate Q_L is

$$Q_L = \frac{W_L}{(3600)(\rho_L)}, \text{ft}^3/\text{s}. \tag{19.504}$$

The bubbling area A_b is

$$A_b = \frac{(Q_V)(R_D) + Q_L}{(T_{SF})(V)(R_D)(\text{Foam fac})}, \text{ft}^2 \tag{19.505}$$

Table 19.48 Values of G and H for each tray spacing.

Tray spacing (inches)	G	H
12	0.77174	-0.02964
18	0.93655	-0.02310
24	T_{SF} = 1 for all values of X	
30	0.98057	0.03220
	T_{SF} = 1.0 X < 1.85	
36	0.96583	0.06162
	T_{SF} = 1 when X < 1.85	

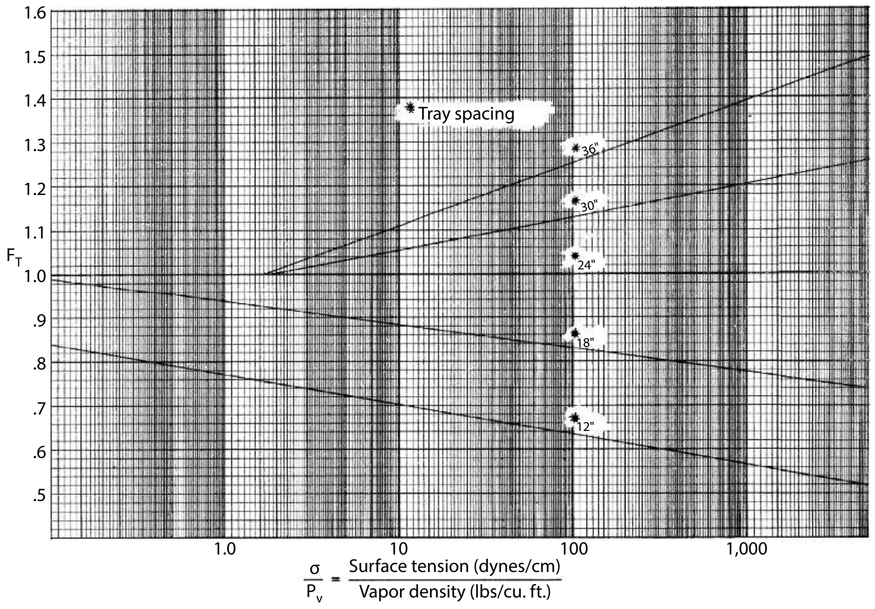


Figure 19.130 Plots of tray spacing vs. ratio of surface tension to vapor density. By permission of Nutter Engineering.

Note: If $(V)(R_D) > 0.5$, then replace 0.5 for $(V)(R_D)$ in Equation (19.505). Table 19.49 gives recommended values of the foam factor and residence time.

Using Table 19.49 to obtain both the foam factor (Foam fac) and the downcomer residence time τ, the downcomer area A_{DC} is

$$A_{DC} = \frac{(\tau)(Q_L)(12)}{(\text{Foam fac})(T_s)}, \text{ft}^2 \tag{19.506}$$

The expression for the safety factor S_{FAC} is

$$S_{FAC} = K(X)^M \tag{19.507}$$

Equation (19.507) is valid for $3.0 \le X \le 50{,}000$.

where

$K = 0.91146$
$M = -0.03821$

The tower area A_T is

$$A_T = \frac{2(A_{DC}) + A_b}{S_{FAC}}, \text{ft}^2 \tag{19.508}$$

Table 19.50 shows the tower diameter versus its cross-sectional areas, and Tables 19.51 and 19.52 give the recommended downcomer area and bubbling area distribution ratios.

Table 19.49 Recommended values for foam factor and residence time.

Fractionators	**Foam factor**	**Downcomer res. time, sec.**
Straight Run		
Hydrocarbon	1.0	4.0
Low MW Alcohols	1.0	3.5
Crude Fractionator	1.0	4.5
Special		
Rich Oil (Top) Demeth. or Deeth.	0.85	4.5
Rich Oil (BTM) Demeth. or Deeth.	1.0	4.5
Hydrocarbon Still (Top)	1.0	4.0
Hydrocarbon Still (BTM)	1.0	4.5
MEA-DEA Still	0.85	4.5
Glycol-DGA Still	0.8	4.5
Sulfinol Still	1.0	5.0
H_2S Stripper	0.9	4.0
Sour Water Stripper	0.5-0.7	4.5
O_2 Stripper	1.0	3.0
Refrigerated Demeth. or Deeth. (Top)	0.8	8.0
Refrigerated Demeth. or Deeth. (BTM)	1.0	5.0
Absorbers		
Oil-Ambient Temp. (Above 0°F)	0.85	4.5
Oil-Low Temp. (Below 0°F)	0.95	4.5
DGA-DEA-MEA (Contactor)	0.75	4.5
Glycol Contactor	0.65	5.0
Sulfinol Contactor	1.0	5.0
Vacuum Towers		
Crude Vacuum	1.0	5.0

By Permission of Nutter Engineering.

Table 19.50 Tower diameter vs. cross-sectional areas.

Tower diameter		Area	Tower diameter		Area	Tower diameter		Area
Feat	**Inches**	**Sq. Ft.**	**Feat**	**Inches**	**Sq. Ft.**	**Feat**	**Inches**	**Sq. Ft.**
2′-0″	24″	3.142	15′-0″	180″	176.715	28′-0″	336	615.752
2′-6″	30″	4.909	15′-6″	186″	188.692	28′-6″	342	637.940
3′-6″	36″	7.069	16′-0″	192″	201.062	29′-0″	348	660.520
3′-6″	42″	9.621	16′-6″	198″	213.825	29′-6″	354	683.493
4′-0″	48″	12.566	17′-0″	204″	226.980	30′-0″	360	706.858
4′-6″	54″	15.904	17′-6″	210″	240.528	30′-6″	366	730.617
5′-0″	60″	19.635	18′-0″	216″	254.469	31′-0″	372	754.768
5′-6″	66″	23.758	18′-6″	222″	268.803	31′-6″	378	779.311
6′-0″	72″	28.274	19′-0″	228″	283.529	32′-0″	384	804.248
6′-6″	78″	33.183	19′-6″	234″	298.648	32′-6″	390	829.577
7′-0″	84″	38.485	20′-0″	240″	314.159	33′-0″	396	855.299
7′-6″	90″	44.179	20′-6″	246″	330.064	33′-6″	402	881.413
8′-0″	96″	50.266	21′-0″	252″	346.361	34′-0″	408	907.920
8′-6″	102″	56.745	21′-6″	258″	363.050	34′-6″	414	934.820
9′-0″	108″	63.617	22′-0″	264″	380.133	35′-0″	420	962.113
9′-6″	114″	70.882	22′-6″	270″	397.608	35′-6″	426	989.798
10′-0″	120″	78.540	23′-0″	276″	415.476	36′-0″	432	1017.876
10′-6″	126″	86.590	23′-6″	282″	433.736	36′-6″	438	1046.347
11′-0″	132″	95.033	24′-0″	288″	452.389	37′-0″	444	1075.210
11′-6″	138″	103.869	24′-6″	294″	471.435	37′-6″	450	1104.466
12′-0″	144″	113.098	25′-0″	300	490.874	38′-0″	456	1134.115
12′-6″	150″	122.719	25′-6″	306	510.705	38′-6″	462	1164.156
13′-0″	156″	132.733	26′-0″	312	530.929	39′-0″	468	1194.591
13′-6″	162″	143.139	26′-6″	318	551.546	39′-6″	476	1225.417
14′-0″	168″	153.938	27′-0″	324	572.555			
14′-6″	174″	165.130	27′-6″	330	593.957			

By Permission of Nutter Engineer.

Table 19.51 Downcomer dimensions.

H/DIA	L/DIA	A_d/A_t
.0000	.0000	.0000
.0005	.0447	.0000
.0010	.0632	.0001
.0015	.0774	.0001
.0020	.0894	.0002
.0025	.0999	.0002
.0030	.1094	.0003
.0035	.1181	.0004
.0040	.1262	.0004
.0045	.1339	.0005
.0050	.1411	.0006
.0055	.1479	.0007
.0060	.1545	.0008
.0065	.1607	.0009
.0070	.1667	.0010
.0075	.1726	.0011
.0080	.1782	.0012
.0085	.1836	.0013
.0090	.1889	.0014
.0095	.1940	.0016
.0100	.1990	.0017
.0105	.2039	.0018
.0110	.2086	.0020
.0115	.2132	.0021
.0120	.2178	.0022
.0125	.2222	.0024
.0130	.2265	.0025
.0135	.2308	.0027
.0140	.2350	.0028
.0145	.2391	.0030
.0150	.2431	.0031
.0155	.2471	.0033
.0160	.2510	.0034
.0165	.2548	.0036
.0170	.2585	.0037
.0175	.2622	.0039
.0180	.2659	.0041
.0185	.2695	.0042
.0190	.2730	.0044
.0195	.2765	.0046
.0200	.2800	.0048
.0205	.2834	.0050
.0210	.2868	.0051
.0215	.2901	.0053
.0220	.2934	.0055
.0225	.2966	.0057
.0230	.2998	.0059
.0235	.3030	.0061
.0240	.3061	.0063
.0245	.3092	.0065
.0250	.3122	.0067
.0255	.3153	.0069
.0260	.3183	.0071
.0265	.3212	.0073
.0270	.3242	.0075
.0275	.3271	.0077
.0280	.3299	.0079
.0285	.3328	.0081
.0290	.3356	.0083
.0295	.3384	.0085
.0300	.3412	.0087
.0305	.3439	.0090
.0310	.3466	.0092
.0315	.3493	.0094
.0320	.3520	.0096
.0325	.3546	.0098
.0330	.3573	.0101
.0335	.3599	.0103
.0340	.3625	.0105
.0345	.3650	.0108
.0350	.3676	.0110
.0355	.3701	.0112
.0360	.3726	.0115
.0365	.3751	.0117
.0370	.3775	.0119
.0375	.3800	.0122
.0380	.3824	.0124
.0385	.3848	.0127
.0390	.3872	.0129
.0395	.3896	.0132
.0400	.3919	.0134
.0405	.3943	.0137
.0410	.3966	.0139
.0415	.3989	.0142
.0420	.4012	.0144
.0425	.4035	.0147
.0430	.4057	.0149
.0435	.4080	.0152
.0440	.4102	.0155
.0445	.4124	.0157
.0450	.4146	.0160
.0455	.4168	.0162
.0460	.4190	.0165
.0465	.4211	.0168
.0470	.4233	.0171
.0475	.4254	.0173
.0480	.4275	.0176
.0485	.4296	.0179
.0490	.4317	.0181
.0495	.4338	.0184
.0500	.4359	.0187
.0505	.4379	.0190
.0510	.4400	.0193
.0515	.4420	.0195
.0520	.4441	.0198
.0525	.4461	.0201
.0530	.4481	.0204
.0535	.4501	.0207
.0540	.4520	.0210
.0545	.4540	.0212
.0550	.4560	.0215
.0555	.4579	.0218
.0560	.4598	.0221
.0565	.4618	.0224
.0570	.4637	.0227
.0575	.4656	.0230
.0580	.4675	.0233
.0585	.4694	.0236
.0590	.4712	.0230
.0595	.4731	.0242
.0600	.4750	.0245
.0605	.4768	.0248
.0610	.4787	.0251
.0615	.4805	.0254
.0620	.4823	.0257
.0625	.4841	.0260
.0630	.4859	.0263
.0635	.4877	.0266
.0640	.4895	.0270
.0645	.4913	.0273
.0650	.4931	.0276
.0655	.4948	.0279
.0660	.4966	.0282
.0665	.4983	.0285
.0670	.5000	.0288
.0675	.5018	.0292
.0680	.5035	.0295
.0685	.5052	.0298
.0690	.5069	.0301
.0695	.5086	.0304
.0700	.5103	.0308
.0705	.5120	.0311
.0710	.5136	.0314
.0715	.5153	.0318
.0720	.5170	.0321
.0725	.5186	.0324
.0730	.5203	.0327
.0735	.5219	.0331
.0740	.5235	.0334
.0745	.5252	.0337
.0750	.5268	.0341
.0755	.5284	.0344
.0760	.5300	.0347
.0765	.5316	.0351
.0770	.5332	.0354
.0775	.5348	.0358
.0780	.5363	.0361
.0785	.5379	.0364
.0790	.5395	.0368
.0795	.5410	.0371
.0800	.5426	.0375
.0805	.5441	.0378
.0810	.5457	.0382
.0815	.5472	.0385
.0820	.5487	.0389
.0825	.5502	.0392
.0830	.5518	.0396
.0835	.5533	.0399
.0840	.5548	.0403
.0845	.5563	.0406
.0850	.5578	.0410
.0855	.5592	.0413
.0860	.5607	.0417
.0865	.5622	.0421
.0870	.5637	.0424
.0875	.5651	.0428
.0880	.5666	.0431
.0885	.5680	.0435
.0890	.5695	.0439
.0895	.5709	.0442
.0900	.5724	.0446
.0905	.5738	.0449
.0910	.5752	.0453
.0915	.5766	.0457
.0920	.5781	.0460
.0925	.5795	.0464
.0930	.5809	.0468
.0935	.5823	.0472
.0940	.5837	.0475
.0945	.5850	.0479
.0950	.5864	.0483
.0955	.5878	.0486
.0960	.5892	.0490
.0965	.5906	.0494
.0970	.5919	.0498
.0975	.5933	.0501
.0980	.5946	.0505
.0985	.5960	.0509
.0990	.5973	.0513
.0995	.5987	.0517
.1000	.6000	.0520
.1005	.6013	.0524
.1010	.6027	.0528
.1015	.6040	.0532
.1020	.6053	.5036
.1025	.6066	.0540
.1030	.6079	.0544
.1035	.6092	.0547
.1040	.6105	.0551
.1045	.6118	.0555
.1050	.6131	.0559
.1055	.6144	.0563
.1060	.6157	.0567
.1065	.6170	.0571
.1070	.6182	.0575
.1075	.6195	.0579
.1080	.6208	.0583
.1085	.6220	.0583
.1090	.6233	.0591
.1095	.6245	.0595
.1100	.6258	.0598
.1105	.6270	.0602
.1110	.6283	.0606
.1115	.6295	.0610
.1120	.6307	.0614
.1125	.6320	.0619
.1130	.6332	.0623
.1135	.6344	.0627
.1140	.6356	.0631
.1145	.6368	.0635
.1150	.6380	.0639
.1155	.6392	.0643
.1160	.6404	.0647
.1165	.6416	.0651
.1170	.6428	.0655
.1175	.6440	.0659
.1180	.6452	.0663
.1185	.6464	.0667
.1190	.6476	.0671
.1195	.6488	.0676
.1200	.6499	.0680
.1205	.6511	.0684
.1210	.6523	.0688
.1215	.6534	.0692
.1220	.6546	.0696
.1225	.6557	.0701
.1230	.6569	.0705
.1235	.6580	.0709
.1240	.6592	.0713
.1245	.6603	.0717
.1250	.6614	.0721
.1255	.6626	.0726
.1260	.6637	.0730
.1265	.6648	.0734
.1270	.6659	.0738
.1275	.6671	.0743
.1280	.6682	.0747
.1285	.6693	.0751
.1290	.6704	.0755
.1295	.6715	.0760
.1300	.6726	.0764
.1305	.6737	.0768
.1310	.6748	.0773
.1315	.6759	.0777
.1320	.6770	.0781
.1325	.6781	.0785
.1330	.6791	.0790
.1335	.6802	.0794
.1340	.6813	.0798
.1345	.6824	.0803
.1350	.6834	.0807
.1355	.6845	.0811
.1360	.6856	.0816
.1365	.6866	.0820
.1370	.6877	.0825
.1375	.6887	.0829
.1380	.6898	.0833
.1385	.6908	.0838
.1390	.6919	.0842
.1395	.6929	.0847
.1400	.6940	.0851
.1405	.6950	.0855
.1410	.6960	.0860
.1415	.6971	.0864
.1420	.6981	.0869
.1425	.6991	.0873
.1430	.7001	.0878
.1435	.7012	.0882
.1440	.7022	.0886
.1445	.7032	.0891
.1450	.7042	.0895
.1455	.7052	.0900
.1460	.7062	.0904
.1465	.7072	.0909
.1470	.7082	.0913
.1475	.7092	.0918
.1480	.7102	.0922
.1485	.7112	.0927
.1490	.7122	.0932
.1495	.7132	.0936
.1500	.7141	.0941
.1505	.7151	.0945
.1510	.7161	.0950
.1515	.7171	.0954
.1520	.7180	.0959
.1525	.7190	.9063
.1530	.7200	.0968
.1535	.7209	.0973
.1540	.7219	.0977
.1545	.7229	.0982
.1550	.7238	.0986
.1555	.7248	.0991
.1560	.7257	.0996
.1565	.7267	.1000
.1570	.7276	.1005
.1575	.7285	.1009
.1580	.7295	.1014
.1585	.7304	.1019
.1590	.7314	.1023
.1595	.7323	.1028
.1600	.7332	.1033
.1605	7341	1037
.1610	.7351	.1042
.1615	.7360	.1047
.1620	.7369	.1051
.1625	.7378	.1056
.1630	.7387	.1061
.1635	.7396	.1066
.1640	.7406	.1070
.1645	.7415	.1075
.1650	.7424	.1080
.1655	.7433	.1084
.1660	.7442	.1089
.1665	.7451	.1094
.1670	.7460	.1099
.1675	.7468	.1103
.1680	.7477	.1108
.1685	.7486	.1113
.1690	.7495	.1118
.1695	.7504	.1122
.1700	.7513	.1127
.1705	.7521	.1132
.1710	.7530	.1137
.1715	.7539	.1142
.1720	.7548	.1146
.1725	.7556	.1151
.1730	.7565	.1156
.1735	.7574	.1161
.1740	.7582	.1166
.1745	.7591	.1171

(*Continued*)

Table 19.51 Downcomer dimensions. (*Continued*)

H/DIA	L/DIA	A_d/A_t	H/DIA	L/DIA	A_d/A_t	H/DIA	L/DIA	A_d/A_t	H/DIA	L/DIA	A_d/A_t	H/DIA	L/DIA	A_d/A_t
.1750	.7599	.1175	.2095	.8139	.1521	.2715	.8895	.2195	.3065	.9221	.2599	.3420	.9488	.3022
.1755	.7608	.1180	.2100	.8146	.1527	.2720	.8900	.2201	.3070	.9225	.2605	.3425	.9491	.3028
.1760	.7616	.1185	.2105	.8153	.1532	.2725	.8905	.2207	.3075	.9229	.2611	.3430	.9494	.3034
.1765	.7625	.1190	.2110	.8160	.1537	.2730	.8910	.2212	.3080	.9233	.2617	.3435	.9498	.3040
.1770	.7633	.1195	.2115	.8167	.1542	.2735	.8915	.2218	.3085	.9237	.2623	.3440	.9501	.3046
.1775	.7642	.1200	.2120	.8174	.1547	.2740	.8920	.2224	.3090	.9242	.2629	.3445	.9504	.3053
.1780	.7650	.1204	.2125	.8182	.1553	.2745	.8925	.2229	.3095	.9246	.2635	.3450	.9507	.3059
.1785	.7659	.1209	.2130	.8189	.1558	.2750	.8930	.2235	.3100	.9250	.2640	.3455	.9511	.3065
.1790	.7667	.1214	.2135	.8196	.1563	.2755	.8935	.2241	.3105	.9254	.2646	.3460	.9514	.3071
.1795	.7675	.1219	.2140	.8203	.1568	.2760	.8940	.2246	.3110	.9258	.2652	.3465	.9517	.3077
.1800	.7684	.1224	.2145	.8210	.1573	.2765	.8945	.2252	.3115	.9262	.2658	.3470	.9520	.3083
.1805	.7692	.1229	.2150	.8216	.1579	.2770	.8950	.2258	.3125	.9270	.2670	.3475	.9524	.3089
.1810	.7700	.1234	.2155	.8223	.1584	.2775	.8955	.2264	.3130	.9274	.2676	.3480	.9527	.3095
.1815	.7709	.1239	.2160	.8230	.1589	.2780	.8960	.2269	.3135	.9278	.2682	.3485	.9530	.3101
.1820	.7717	.1244	.2165	.8237	.1594	.2785	.8965	.2275	.3140	.9282	.2688	.3490	.9533	.3107
.1825	.7725	.1249	.2170	.8244	.1600	.2790	.8970	.2281	.3145	.9286	.2693	.3495	.9536	.3113
.1830	.7733	.1253	.2175	.8251	.1605	.2795	.8975	.2286	.3150	.9290	.2699	.3119	.3500	.9539
.1835	.7742	.1258	.2180	.8258	.1610	.2800	.8980	.2292	.3155	.9294	.2705	.3505	.9543	.3125
.1840	.7750	.1263	.2185	.8265	.1615	.2805	.8985	.2298	.3160	.9298	.2711	.3510	.9546	.3131
.1845	.7758	.1268	.2190	.8271	.1621	.2810	.8990	.2304	.3165	.9302	.2717	.3515	.9549	.3137
.1850	.7766	.1273	.2195	.8278	.1626	.2815	.8995	.2309	.3170	.9306	.2723	.3520	.9552	.3143
.1855	.7774	.1278	.2200	.8285	.1631	.2820	.8999	.2315	.3175	.9310	.2729	.3525	.9555	.3150
.1860	.7782	.1283	.2205	.8292	.1636	.2825	.9004	.2321	.3180	.9314	.2735	.3530	.9558	.3156
.1865	.7790	.1288	.2210	.8298	.1642	.2830	.9009	.2326	.3185	.9318	.2741	.3535	.9561	.3162
.1870	.7798	.1293	.2215	.8305	.1647	.2835	.9014	.2332	.3190	.9322	.2747	.3540	.9564	.3168
.1875	.7806	.1298	.2220	.8312	.1652	.2840	.9019	.2338	.3195	.9326	.2753	.3545	.9567	.3174
.1880	.7814	.1303	.2225	.8319	.1658	.2850	.9028	.2349	.3200	.9330	.2759	.3550	.9570	.3180
.1885	.7822	.1308	.2230	.8325	.1663	.2855	.9033	.2355	.3205	.9333	.2765	.3555	.9573	.3186
.1890	.7830	.1313	.2235	.8332	.1668	.2860	.9038	.2361	.3210	.9337	.2771	.3560	.9576	.3192
.1895	.7838	.1318	.2240	.8338	.1674	.2865	.9043	.2367	.3215	.9341	.2777	.3565	.9579	.3198
.1900	.7846	.1323	.2245	.8345	.1679	.2870	.9047	.2872	.3220	.9345	.2782	.3570	.9582	.3204
.1905	.7854	.1328	.2250	.8352	.1684	.2875	.9052	.2378	.3225	.9349	.2788	.3575	.9585	.3211
.1910	.7862	.1333	.2255	.8358	.1689	.2880	.9057	.2384	.3230	.9352	.2794	.3580	.9588	.3217
.1915	.7870	.1338	.2260	.8365	.1695	.2885	.9061	.2390	.3235	.9356	.2800	.3585	.9591	.3223
.1920	.7877	.1343	.2265	.8371	.1700	.2890	.9066	.2395	.3240	.9360	.2806	.3590	.9594	.3229
.1925	.7885	.1348	.2270	.8378	.1705	.2895	.9071	.2401	.3245	.9364	.2812	.3595	.9597	.3235
.1930	.7893	.1353	.2275	.8384	.1711	.2900	.9075	.2407	.3250	.9367	.2818	.3600	.9600	.3241
.1935	.7901	.1358	.2280	.8391	.1716	.2905	.9080	.2413	.3255	.9371	.2824	.3605	.9603	.3247
.1940	.7909	.1363	.2285	.8397	.1721	.2910	.9084	.2419	.3260	.9375	.2830	.3610	.9606	.3253
.1945	.7916	.1368	.2290	.8404	.1727	.2915	.9089	.2424	.3265	.9379	.2511	.3615	.9609	.3259
.1950	.7924	.1373	.2295	.8410	.1732	.2920	.9094	.2430	.3270	.9382	.2842	.3620	.9612	.3265
.1955	.7932	.1378	.2300	.8417	.1738	.2925	.9098	.2436	.3275	.9386	.2848	.3625	.9614	.3272
.1960	.7939	.1383	.2305	.8423	.1743	.2930	.9103	.2442	.3280	.9390	.2854	.3630	.9617	.3278
.1965	.7947	.1388	.2310	.8429	.1748	.2935	.9107	.2448	.3285	.9393	.2860	.3635	.9620	.3284
.1970	.7955	.1393	.2315	.8436	.1754	.2940	.9112	.2453	.3290	.9397	.2866	.3640	.9623	.3290
.1975	.7962	.1398	.2320	.8442	.1759	.2945	.9116	.2459	.3295	.9401	.2872	.3645	.9626	.3296
.1980	.7970	.1403	.2325	.8449	.1764	.2950	.9121	.2465	.3300	.9404	.2878	.3650	.9629	3302
.1985	.7977	.1409	.2330	.8455	.1770	.2955	.9125	.2471	.3305	.9408	.2884	.3655	.9631	.3308
.1990	.7985	.1414	.2335	.8461	.1775	.2960	.9130	.2477	.3310	.9411	.2890	.3660	.9634	.3315
.1995	.7992	.1419	.2340	.8467	.1781	.2965	.9134	.2482	.3315	.9415	.2896	.3665	.9637	.3321
.2000	.8000	.1424	.2345	.8474	.1786	.2970	.9139	.2488	.3320	.9419	.2902	.3675	.9642	.3333
.2005	.8007	.1429	.2350	.8480	.1791	.2975	.9143	.2494	.3325	.9422	.2908	.3680	.9645	.3339
.2010	.8015	.1434	.2355	.8486	.1797	.2980	.9148	.2500	.3330	.9426	.2914	.3685	.9648	.3345
.2015	.8022	.1439	.2360	.8492	.1802	.2985	.9152	.2506	.3335	.9429	.2920	.3690	.9651	.3351
.2020	.8030	.1444	.2365	.8499	.1808	.2990	.9156	.2511	.3340	.9433	.2926	.3695	.9653	.3357
.2025	.8037	.1449	.2370	.8505	.1813	.2995	.9161	.2517	.3345	.9436	.2932	.3700	.9656	.3364
.2030	.8045	.1454	.2650	.8827	.2122	.3000	.9165	.2523	.3350	.9440	.2938	.3705	.9659	.3370
.2035	.8052	.1460	.2655	.8832	.2128	.3005	.9170	.2529	.3335	.9443	.2944	.3710	.9661	.3376
.2040	.8059	.1465	.2660	.8837	.2133	.3010	.9174	.2535	.3360	.9447	.2950	.3715	.9664	.3382
.2045	.8067	.1470	.2665	.8843	.2139	.3015	.9178	.2541	.3365	.9450	.2956	.3720	.9667	.3388
.2050	.8074	.1475	.2670	.8848	.2145	.3020	.9138	.2547	.3370	.9454	.2962	.3750	.9669	.3394
.2055	.8081	.1480	.2675	.8853	.2150	.3025	.9187	.2552	.3375	.9457	.2968	.3730	.9672	.3401
.2060	.8089	.1485	.2680	.8858	.2156	.3030	.9191	.2558	.3380	.9461	.2974	.3735	.9676	.3407
.2065	.8096	.1490	.2685	.8864	.2161	.3035	.9195	.2564	.3385	.9464	.2980	.3740	.9677	.3413
.2070	.8103	.1496	.2690	.8869	.2167	.3040	.9200	.2570	.3390	.9467	.2986	.3745	.9680	.3419
.2075	.8110	.1501	.2695	.8874	.2173	.3045	.9204	.2576	.3400	.9474	.2998			
.2080	.8118	.1506	.2700	.8879	.2178	.3050	.9208	.2582	.3405	.9478	.3004			
.2085	.8125	.1511	.2705	.8884	.2184	.3055	.9212	.2588	.3410	.9481	.3010			
.2090	.8132	.1516	.2710	.8890	.2190	.3060	.9217	.2593	.3415	.9484	.3016			

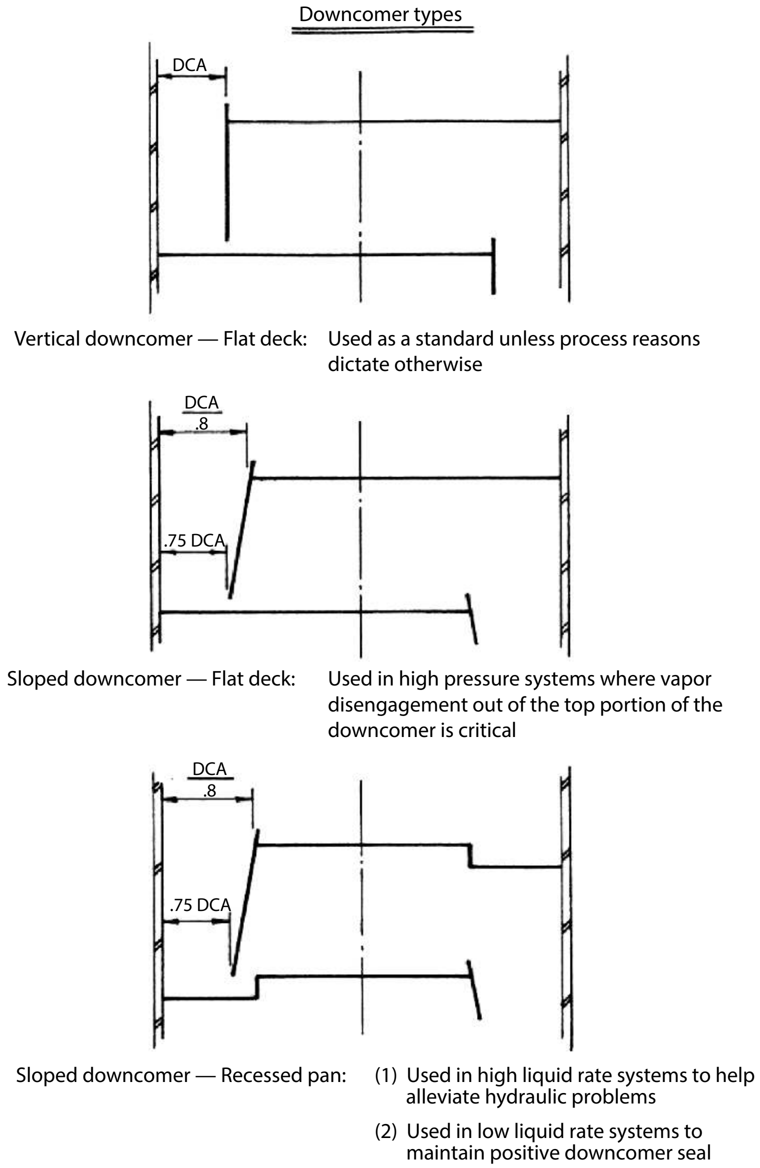
By Permission of Nutter Engineer.

Table 19.52 Downcomer dimensions.

H/DIA	L/DIA	A_d/A_t	H/DIA	L/DIA	A_d/A_t	H/DIA	L/DIA	A_d/A_t	H/DIA	L/DIA	A_d/A_t	H/DIA	L/DIA	A_d/A_t
.3750	.9682	.3425	.4005	.9800	.3742	.4260	.9890	.4061	.4515	.9953	.4383	.4770	.9989	.4707
.3755	.9685	.3431	.4010	.9802	.3748	.4265	.9891	.4068	.4520	.9954	.4390	.4775	.9990	.4714
.3760	.9688	.3438	.4015	.9804	.3754	.4270	.9893	.4074	.4525	.9955	.4396	.4780	.9990	.4720
.3765	.9690	.3444	.4020	.9806	.3760	.4275	.9894	.4080	.4530	.9956	.4402	.4785	.9991	.4726
.3770	.9693	.3453	.4025	.9808	.3467	.4280	.9896	.4086	.4535	.9957	.4409	.4790	.9991	.4733
.3775	.9695	.3456	.4030	.9810	.3773	.4285	.9897	.4093	.4540	.9958	.4415	.4795	.9992	.4739
.3780	.9698	.3462	.4035	.9812	.3779	.4290	.9899	.4099	.4545	.9959	.4421	.4800	.9992	.4745
.3785	.9700	.3468	.4040	.9814	.3785	.4295	.9900	.4105	.4550	.9959	.4428	.4805	.9992	.4752
.3790	.9703	.3475	.4045	.9816	.3791	.4300	.9902	.4112	.4555	.9960	.4434	.4810	.9993	.4758
.3795	.9705	.3481	.4050	.9818	.3798	.4305	.9903	.4118	.4560	.9961	.4440	.4815	.9993	.4765
.3800	.0708	.3487	.4055	.9802	.3804	.4310	.9904	.4124	.4565	.9962	.4447	.4820	.9994	.4771
.3805	.9710	.3493	.4060	.9822	.3810	.4315	.9906	.4131	.4570	.9963	.4453	.4825	.9994	.4777
.3810	.9713	.3499	.4065	.9824	.3816	.4320	.9907	.4137	.4575	.9964	.4460	.4830	.9994	.4784
.3815	.9715	.3505	.4070	.9825	.3823	.4325	.9908	.4143	.4580	.9965	.4466	.4835	.9995	.4790
.3820	.9718	.3512	.4075	.9827	.3829	.4330	.9910	.4149	.4585	.9965	.4472	.4840	.9995	.4796
.3825	.9720	.3518	.4080	.9829	.3835	.4335	.9911	.4156	.4590	.9966	.4479	.4845	.9995	.4803
.3830	.9722	.3524	.4085	.9831	.3842	.4340	.9912	.4162	.4595	.9967	.4485	.4850	.9995	.4809
.3835	.9725	.3530	.4090	.9833	.3848	.4345	.9914	.4168	.4600	.9968	.4491	.4855	.9996	.4815
.3840	.9727	.3536	.4095	.9835	.3854	.4350	.9915	.4175	.4605	.9969	.4498	.4860	.9996	.4822
.3845	.9730	.3543	.4100	.9837	.3860	.4355	.9916	.4181	.4610	.9970	.4505	.4865	.9996	.4828
.3850	.9732	.3549	.4105	.9838	.3867	.4360	.9918	.4187	.4615	.9970	.4510	.4870	.9997	.4834
.3855	.9734	.3555	.4110	.9840	.3873	.4365	.9919	.4194	.4620	.9971	.4517	.4875	.9997	.4841
.3860	.9737	.3561	.4115	.9842	.3879	.4370	.9920	.4200	.4625	.9972	.4523	.4880	.9997	.4847
.3865	.9739	.3567	.4120	.9844	.3885	.4375	.9922	.4206	.4630	.9973	.4529	.4885	.9997	.4854
.3870	.9741	.3574	.4125	.9846	.3892	.4380	.9923	.4213	.4635	.9973	.4536	.4890	.9998	.4860
.3875	.9744	.3580	.4130	.9847	.3898	.4385	.9924	.4219	.4640	.9974	.4542	.4895	.9998	.4866
.3880	.9746	.3586	.4135	.9849	.3904	.4390	.9925	.4225	.4645	.9975	.4548	.4900	.9998	.4873
.3885	.9748	.3592	.4140	.9851	.3910	.4395	.9927	.4232	.4650	.9975	.4555	.4905	.9998	.4879
.3890	.9750	.3598	.4145	.9853	.3917	.4400	.9928	.4238	.4655	.9976	.4561	.4910	.9998	.4885
.3895	.9753	.3605	.4150	.9854	.3923	.4405	.9929	.4244	.4660	.9977	.4567	.4915	.9999	.4892
.3900	.9755	.3611	.4155	.9856	.3929	.4410	.9930	.4251	.4665	.9978	.4574	.4920	.9999	.4898
.3905	.9757	.3617	.4160	.9858	.3936	.4415	.9931	.4257	.4670	.9978	.4580	.4925	.9999	.4905
.3910	.9759	.3623	.4165	.9860	.3942	.4420	.9932	.4263	.4675	.9979	.4568	.4930	.9999	.4911
.3915	.9762	.3629	.4170	.9861	.3948	.4425	.9934	.4270	.4680	.9979	.4593	.4935	.9999	.4917
.3920	.9764	.3636	.4175	.9863	.3642	.4430	.9935	.4276	.4685	.9980	.4599	.4940	.9999	.4924
.3925	.9766	.3642	.4180	.9865	.3961	.4435	.9936	.4282	.4690	.9981	.4606	.4945	.9999	.4930
.3930	.9768	.3648	.4185	.9866	.3967	.4440	.9937	.4288	.4695	.9981	.4612	.4950	1.0000	.4936
.3935	.9771	.3654	.4190	.9868	.3973	.4445	.9938	.4295	.4700	.9982	.4618	.4955	1.0000	.4943
.3940	.9773	.3661	.4195	.9870	.3979	.4450	.9939	.4301	.4705	.9983	.4625	.4960	1.0000	.4949
.3945	.9775	.3667	.4200	.9871	.3986	.4455	.9940	.4307	.4710	.9983	.4631	.4965	1.0000	.4955
.3950	.9777	.3673	.4205	.9873	.3992	.4460	.9942	.4314	.4715	.9984	.4637	.4970	1.0000	.4962
.3955	.9779	.3679	.4210	.9874	.3998	.4465	.9943	.4320	.4720	.9984	.4644	.4975	1.0000	.4968
.3960	.9781	.3685	.4215	.9876	.4005	.4470	.9944	.4326	.4725	.9985	.4650	.4980	1.0000	.4975
.3965	.9783	.3692	.4220	.9878	.4011	.4475	.9945	.4333	.4730	.9985	.4656	.4985	1.0000	.4981
.3970	.9786	.3698	.4225	.9879	.4017	.4480	.9946	.4339	.4735	.9986	.4663	.4990	1.0000	.4987
.3975	.9788	.3704	.4230	.9881	.4023	.4485	.9947	.4345	.4740	.9986	.4669	.4995	1.0000	.4994
.3980	.9790	.3710	.4235	.9882	.4030	.4490	.9948	.4352	.4745	.9987	.4675	.5000	1.0000	.5000
.3985	.9792	.3717	.4240	.9884	.4036	.4495	.9949	.4358	.4750	.9987	.4682			
.3990	.9794	.3723	.4245	.9885	.4042	.4500	.9950	.4364	.4755	.9988	.4688			
.3995	.9796	.3729	.4250	.9887	.4049	.4505	.9951	.4371	.4760	.9988	.4695			
.4000	.9798	.3735	.4255	.9888	.4055	.4510	.9952	.4377	.4765	.9989	.4701			

Table 19.53 Determining the number of tray passes.

Tray spacing (inch)	Increase number of passes if gpm/in. of weir exceeds
12	3
18	18
24	13



Vertical downcomer — Flat deck: Used as a standard unless process reasons dictate otherwise

Sloped downcomer — Flat deck: Used in high pressure systems where vapor disengagement out of the top portion of the downcomer is critical

Sloped downcomer — Recessed pan:
(1) Used in high liquid rate systems to help alleviate hydraulic problems
(2) Used in low liquid rate systems to maintain positive downcomer seal

Figure 19.131 Dimensions of downcomer types. By permission of Nutter Engineering.

The column diameter D is

$$D = \left(\frac{4 A_T}{\pi} \right)^{0.5} \tag{19.509}$$

The number of tray passes required is determined from Table 19.53.

Figures 19.131 and 19.132 show the dimensions of the various downcomers and the distribution area for the number of passes.

19.52.1 Diameter of Sieve/Valve Trays (F Factor)

The superficial vapor velocity above the tray not occupied by downcomers is

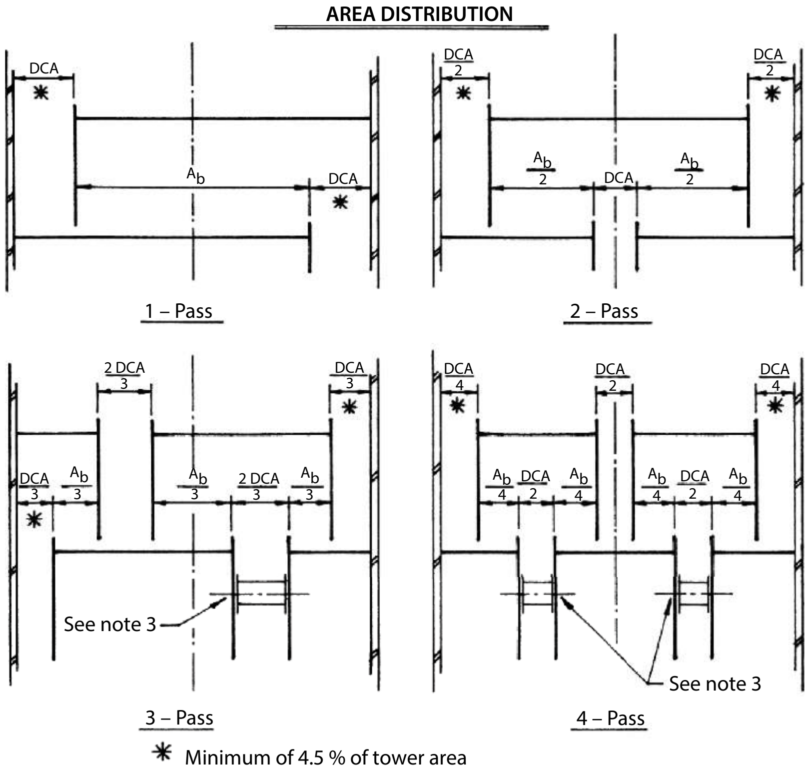


Figure 19.132 Area distribution for different pass types. By permission of Nutter Engineering.

$$U = \frac{F}{\rho_V^{0.5}} \tag{19.510}$$

where F is the factor based on free column cross-sectional area (total column area minus the downcomer area). For foaming systems, the F factor should be multiplied by 0.75. Figure 19.133 shows the F factor correlation as a function of column pressure for various tray spacings [159].

Branan [159] developed an equation for the F factor as follows:

$$F = (547 - 173.2T + 2.3194T^2)10^{-6}P + 0.32 + 0.0847T - 0.00078T^2 \tag{19.511}$$

Correlation ranges are

F =	0.8 to 2.4
F =	0 to 220
F =	18 to 36

For estimating the downcomer area, Branan has developed an empirical equation to fit Frank's Figure 19.134 [159] as

$$D_L = 6.667T + 16.665 \tag{19.512}$$

Clear liquid velocity (ft/s) through the downcomer is then found by multiplying D_L by 0.00223. The correlation is not valid if $(\rho_L - \rho_V) < 30\,\mathrm{lb/ft^3}$ (very high-pressure systems). For foaming systems, D_L should be multiplied by 0.7.

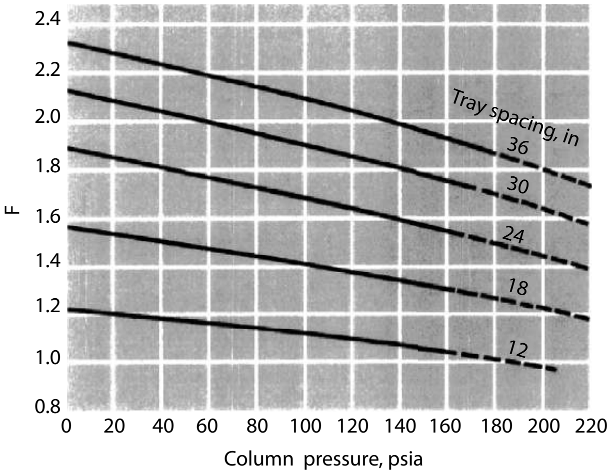
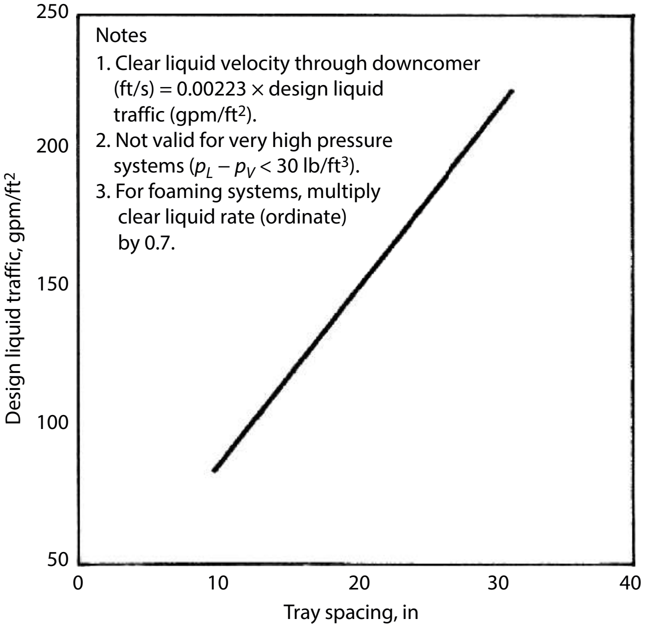


Figure 19.133 F Factor as a function of column pressure drop and tray spacing (Source: Branan, C., "Rules of Thumb for Chemical Engineers", 4th Ed., Gulf Publishing Professional, 2005).



Notes
1. Clear liquid velocity through downcomer (ft/s) = 0.00223 × design liquid traffic (gpm/ft^2).
2. Not valid for very high pressure systems ($p_L - p_V < 30$ lb/ft^3).
3. For foaming systems, multiply clear liquid rate (ordinate) by 0.7.

Figure 19.134 Estimation of downcomer area for a tray-type distillation column (Source: Branan, C., "Rules of Thumb for Chemical Engineers", 4th Ed., Gulf Publishing Professional 2005).

Frank recommends segmental downcomers of at least 5% of total column cross-sectional area, regardless of the area obtained by this correlation. For final design, complete tray hydraulic calculations are required. For even faster estimates, Branan proposed the following rough F factor guidelines.

Situation	**F factor**
Fractionating column total cross section vapor velocity	1.0 – 1.5
Sieve tray hole velocity to avoid weeping	> 12
Disengaging equipment for separating liquid droplets from vapor	< 6

19.52.2 Diameter of Sieve/Valve Trays (Lieberman)

Lieberman [188] provides two rules of thumb for troubleshooting fractionators, which could also be used as checks on a design. The pressure drop (ΔP) across a section of trays must not exceed 22% of the space between the tray decks to avoid incipient flood. This is expressed by

$$\frac{\Delta P}{SG}(T_n)(T_s) < 22\% \tag{19.513}$$

where

ΔP = Pressure drop in inches of water.
SG = Specific gravity of the liquid on the tray at the appropriate temperature.
T_n = Number of trays
T_s = Tray spacing, in.

For sieve trays, a spray height of 15 in. is obtained when the jetting factor is 6–7.

$$\text{Jetting factor} = U^2 \rho_V / \rho_L \tag{19.514}$$

where

U = Hole vapor velocity, ft/s.
ρ_L = Liquid density.
ρ_V = Vapor density.

For a 15-inch spray height, a tray spacing of at least 21 inches is recommended.

Example 19.48: Column Sizing Using Nutter's Valve Tray

Size an absorber tower for Methyl ethyl amine (MEA) system with the following conditions:

Vapor flow, lb/h	40,000
Vapor density, lb/ft³	0.295
Liquid flow, lb/h	330,000
Liquid density, lb/ft³	61.85
Foam factor	0.75
Residence time, s	4.5
Tray spacing, in.	24
Surface tension, dyne/cm.	57.6

Solution

An Excel spreadsheet program (Example 19.48.xlsx) has been developed that determines the parameters required for sizing a tower using valve trays. The program uses Nutter's Float Design Manual and calculates the bubbling area, downcomer area, tower area, and diameter. The tower diameter may be rounded to an appropriate value and the area is recalculated. Once the tower diameter is set, the downcomer area ratio is determined and the downcomer height is computed from Tables 19.51 and 19.52; and Table 19.53 shows the number of tray passes. Table 19.54 shows the results for tower sizing using valve trays from the Excel spreadsheet (Example 19.48.xlsx). The results give the tower area of 26.54 ft^2 and a diameter of 5.81 ft.

Tray Geometry Sizing

From the tower size (normally by increasing the diameter to the nearest 6-inch increment), the following procedure is used to calculate the downcomer dimensions.

The downcomer area ratio is

$$R_{DCA} = \frac{A_{DC}}{(2A_{DC} + A_b)} \tag{19.515}$$

Table 19.54 Results of tower design for valve trays of Example 19.48.

Tower design for valve trays	
Vapor flow rate	40,000 lb/h
Vapor volumetric flow rate	37.665 ft^3/s
Liquid flow rate	330,000 lb/h
Liquid volumetric flow rate	1.482 ft^3/s
Liquid density	61.85 lb/ft^3
Vapor density	0.295 lb/ft^3
Foam factor	0.75
Downcomer residence time	4.5 sec
Tray spacing	24 in.
Surface tension	57.6 dyne/cm
Surface tension/Vapor density	195.25
Safety factor	0.745
Density radical	0.069
Bubbling area	10.882 ft^2
Downcomer area	4.446 ft^2
Tower area	26.542 ft^2
Tower diameter	5.81 ft.
Downcomer area ratio	0.225

The calculated tower diameter is increased to 6.0 ft. and the tower area becomes

$$\begin{aligned} A &= \frac{\pi d^2}{4}, \text{ft}^2 \\ &= \frac{\pi \times 6^2}{4} \\ &= 28.27 \text{ ft}^2 \end{aligned}$$

Using Tables 19.51 and 19.52 and R_{DCA} will allow a direct computation of chord height for the side downcomer. This is rounded up to the nearest ½-in. increment dimension. Using this dimension, the corrected downcomer area is calculated from Tables 19.51 and 19.52. The computed output of the tower used by the downcomer area (R_{DCA}) is 0.2246.

That is,

$$\frac{A_d}{A_t} = 0.2246$$

H/DIA	L/DIA	A_d/A_t
0.2760	0.8940	0.2246

The corresponding downcomer height to tower diameter ratio is 0.2760.

$$\begin{aligned} \text{The downcomer height} &= (0.2760)(6)(12) \\ &= 19.87 \text{ in.} + \tfrac{1}{2} \text{ in.} \\ &= 20.4 \text{ in.} \end{aligned}$$

$$\begin{aligned} \text{The new H/DIA} &= 20.4/72 \\ &= 0.2833 \end{aligned}$$

This gives a downcomer area to tower area ratio of 0.2329.

H/DIA	L/DIA	A_d/A_t
0.2830	0.9009	0.2326
Avg 0.28325	Avg 0.90115	Avg 0.2329
0.2835	0.9014	0.2332

That is,

$$\frac{A_d}{A_t} = 0.2329$$

$$\text{The downcomer area } A_d = (0.2329)(28.27)$$

$$= 6.584\ \text{ft}^2$$

$$\text{The weir height, L} = (0.90115 \times 72)$$
$$= 64.88\ \text{in}\ (65\ \text{in.})$$

The computed liquid flow rate is

$$= \frac{(33{,}0000)}{(61.82)(3600)} \left\{ \frac{\text{lb}}{\text{h}} \cdot \frac{\text{ft}^3}{\text{lb}} \cdot \frac{\text{h}}{\text{s}} \right\}$$
$$= 1.483\ \text{ft}^3/\text{s}\ (665.35\ \text{US gpm.})$$

The liquid flow per inch of weir = 665.35/65 = 10.24 US gpm per inch of weir.

A single-pass tray is selected.

19.53 Troubleshooting, Predictive Maintenance, and Controls for Distillation Columns

Troubleshooting currently is now much more sophisticated due to the technical tools that are available for investigating and analyzing performance. Radiation scanning referred to as distillation column scanning or "gamma scanning" is now in common use. This is the process of studying the process material inside a vessel by placing a moving radioactive source inside it and moving radiation detector along the exterior of the vessel (Figure 19.135a). Such scanning enables the engineer to study tray or packing hydraulics inside the vessel at any set of online conditions as it provides essential data to

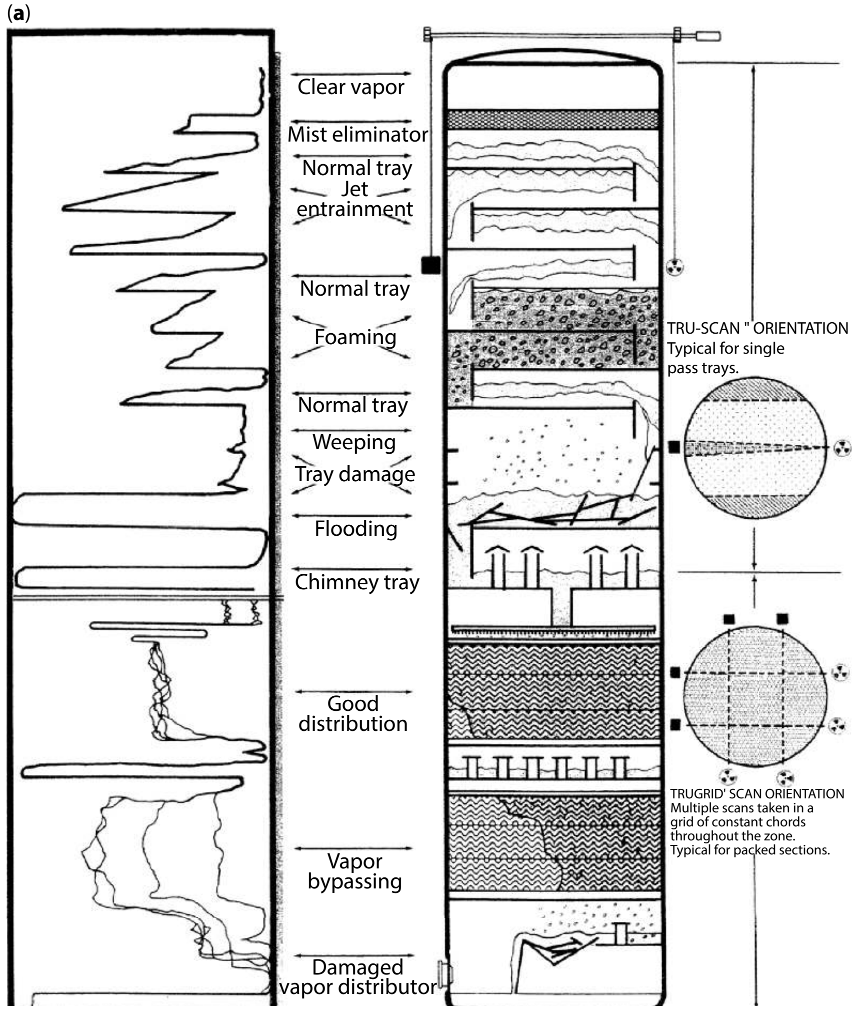
- Track the performance—deteriorating effects of fouling, foaming, flooding, and weeping.
- Optimize the performance of vessels.
- Identify maintenance requirements in advance of scheduled turnarounds.
- Extend vessel run times.

Figure 19.135b shows how the radiation source and the detector are aligned on opposite sides of a column or vessel. There are two typical alignments: one-pass trayed column and a packed tower, respectively. The source and the detector are synchronized and lowered down the column, while the intensity measurements are recorded at specified intervals. The variation in measured intensity depends on the changes in "absorber" thickness and density. The absorber is a composite of the vessel walls, insulation, column internal trays, packings, downcomers, support grids, gas/liquid distributors, and process material inside the column. Thus, the difference in intensity that is detected is the result of process changes such as density differences between liquid and vapor within the column.

The scan produces a density profile that shows the relative density vs. column elevation. The results are then interpreted to show the behavior of the column, its performance, and physical condition. Figure 19.135c illustrates how scan results are used to determine tray loading, extent of flooding, and vapor space conditions. Tray locations and sources of external interference (e.g., welds, stiffening, rings, or supports) are on the left side of the plot.

Flooding: Different extents of flooding occur on tray -1 in the region where scan profile flattens out and rises up.

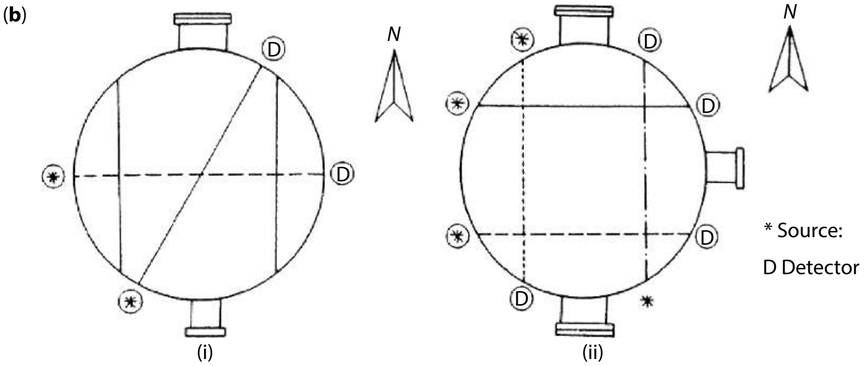
Clear vapor region: Is marked by the bar line on the right side of the plot. It defines statistically what the relative density should be when only vapor is present.



TRU-SCAN " ORIENTATION
Typical for single
pass trays.

TRUGRID' SCAN ORIENTATION
Multiple scans taken in a
grid of constant chords
throughout the zone.
Typical for packed sections.

Figure 19.135a Distillation diagnostics (Source: Tru-Tec Division, Koch Engineering Co. Inc.).



Double scan alignments for a single-pass trayed column. A single radiation souce and a detector face each other along the column diameter in both alignments. They are moved synchronously up and down the column along its other surface.

Alignments for four separate scans of a packed tower. In such cases, scans are completed in parallel and orthogonal 2 x 2 grid. The distance from source to detector must remain the same for all four scan lines. Therefore, the source and the detector are located at ends of each four equidistant chords oriented orthogonally in a north-to-south and east-to-west pattern.

Figure 19.135b Gamma ray scanning of a single-pass tray and of a packed tower.

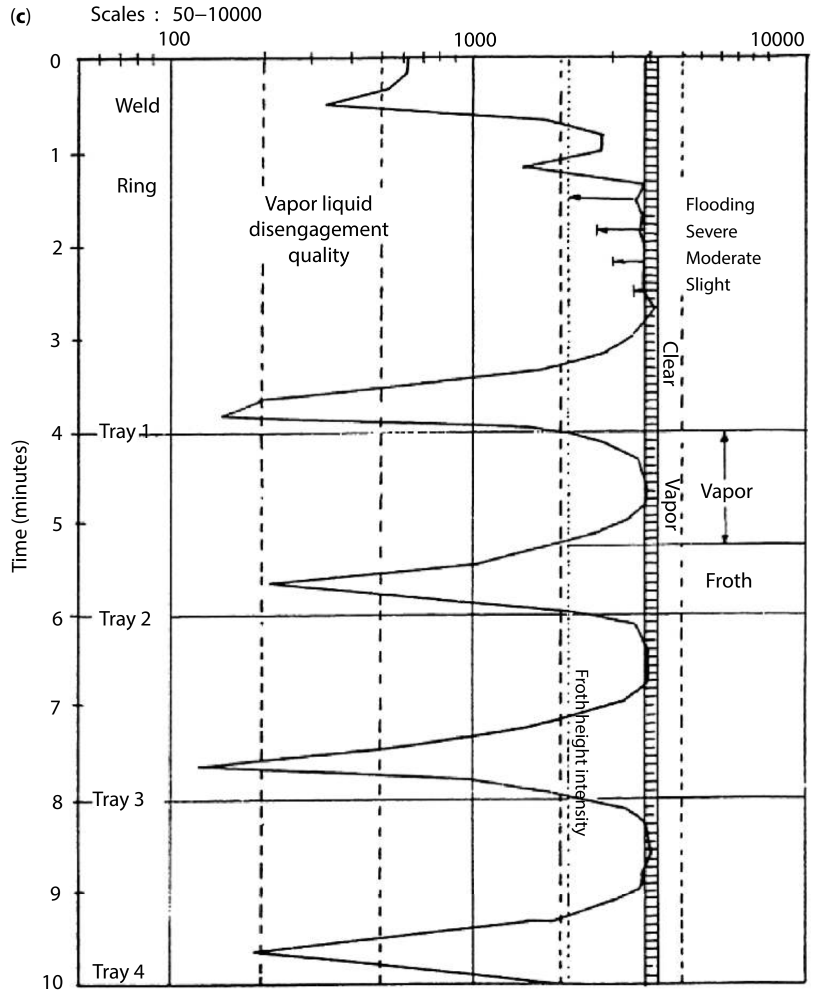


Figure 19.135c Gamma ray scanning profile of a distillation column. The profile is presented as a plot of time (and tray no.) vs. intensity of absorbed emission. The solid line depicts the actual scan data: the clear vapor region is marked by bar line. The dotted vertical lines to the left of the clear vapor bar represent froth height intensity. It is also called datum line (Source: Bowman, J. D., Chem. Eng. Prog., Feb. 1991).

Froth height intensity: The dotted line to the left of the clear vapor bar is labeled as froth height intensity. This is also referred to as the datum line through the statistical average of the relative vapor space densities and the liquid densities appearing on the tray active areas.

Forth height: Is the vertical distance between the deck and spray height level. This is depicted on tray -2 on the right side of the vapor bar (Figure 19.135c). The tray is sustaining approximately 9 in. (229 mm) (± 1 in.) of aerated froth.

Clear vapor height: Is the tray spacing minus froth height.

Experience with the techniques is important for its effective use combining it with a computer model can assist in both theoretical and practical design applications; best results are obtained by combining scan results and tray packing design [189, 190]; it provides data that can aid significantly in determining whether a column is having liquid/vapor flow and or distribution problems. This system provides an accurate density profile of the operating fluids on each tray or through the packing of a packed column, which can be used to identify a number of column malfunctions as shown in Table 19.55.

Figure 19.135d shows an example of using gamma scanning in a depropanizer column to evaluate its performance.

Table 19.55 Column malfunctions identifiable by scanning.

Type of malfunctions	Particulars
Process	Liquid hold up on tray/packing due to fouling or plugging. Foaming Subcooled or superheated feed or reflux
Rate Related	Entrainment – slight, moderate, severe jet flooding. Flooding. Tray starvation. Weeping.
Mechanical	Damage to trays and packings. Tray displacement. Damage to trays due to corrosion. Dislocation of liquid or vapor distributors. Missing tray manways. Level control problems in the base of the column.

Other troubleshooting techniques include computer modeling, checking the reliability of instrumentation, measuring quality of product streams with varying reflux rates, measuring column tray temperatures at close intervals, stabilizing the feed rate, bottoms withdrawal, and overhead condensing rates. Some surprising problems can be identified, including

1. Trays may have damage to caps, valves, distributors, sieve holes, or packing for packed towers.
2. The contacting devices of (1) above may actually be missing, i.e., blown off one or more trays, so all that is existing is a "rain-deck" tray with no liquid–vapor contacting.
3. Crud, polymer, gunk and other processing residues, plus maintenance tools, rags, or overalls may be plugging or corroding the liquid flow paths.
4. Entrainment.
5. Weeping of trays or flooding of packing or trays.
6. Foaming limitations.
7. Unusual feed conditions, unexpected or uncontrolled.
8. Many other situations, almost too odd to imagine. References on this topic include 190–199.

The topic of control of distillation columns has been discussed by many authorities with a wide variety of experience [200–203] and is too specialized to be covered in this text.

Fractionating Tray Stability Diagrams

Tray stability diagrams provide an analytical means to visualize what is occurring on a fractionating tray. Figure 19.136A provides an example of a tray stability diagram. The areas of unacceptable operation are as follows:

- Flooding
- Downcomer Filling
- Blowing
- Weeping/Dumping

Areas of Unacceptable Operation

Other flooding mechanisms are as follows.

(**d**)

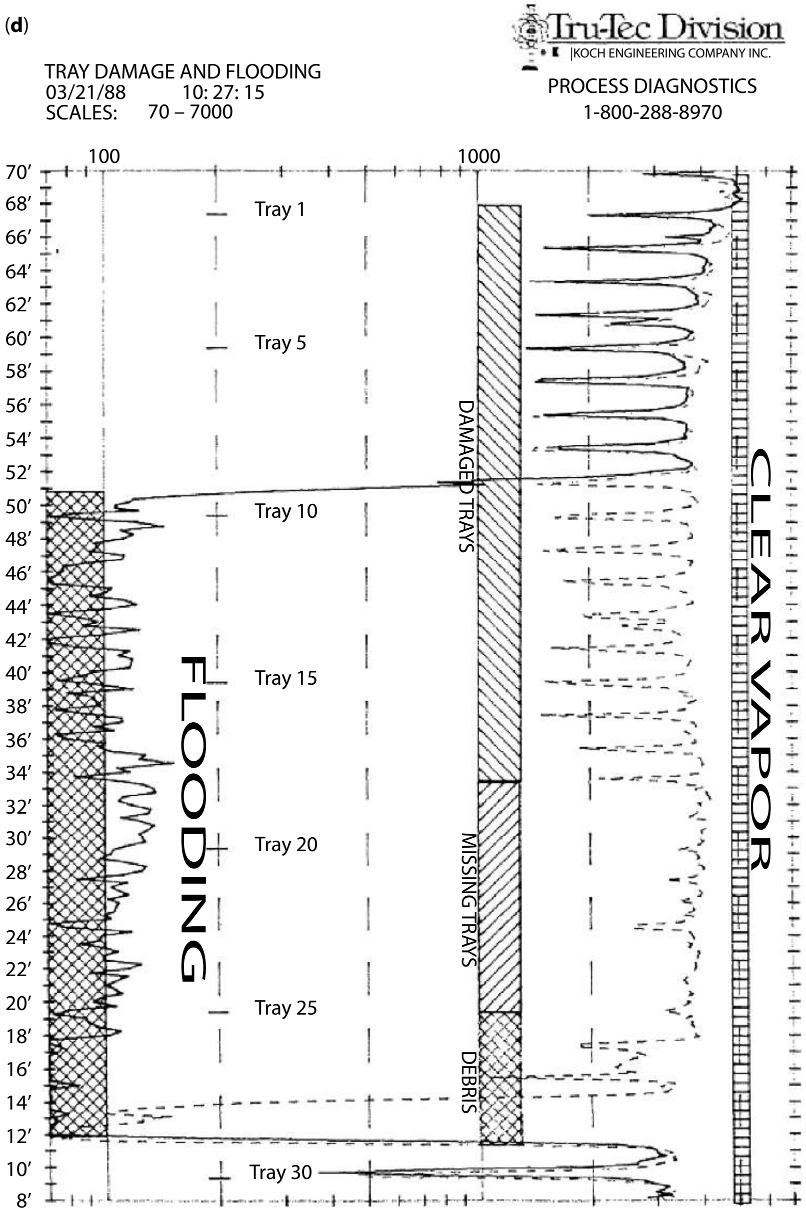


Figure 19.135d Gamma ray scanning "diagnostics" of depropanizer column to evaluate performance. (Used by permission, Tru-Tec Division, Koch Engineering Co. Inc).

Spray Entrainment Flooding

At low liquid flow rates, the spray height increases as the vapor velocity increases. The trays operate in a spray regime, and the spray may eventually reach the downside of the tray above, causing a substantial carryover of the liquid droplets. As the vapor velocity is increased, a point is reached where a considerable amount of the liquid droplets is carried (entrained) to the tray above along with the vapor.

Froth Entrainment Flooding

Froth appears on a tray over a range of gas and liquid flow rates. Depending upon the flow rates of the phase and the tray spacing, the froth may almost fill the entire space between two successive trays. At higher liquid loads, when

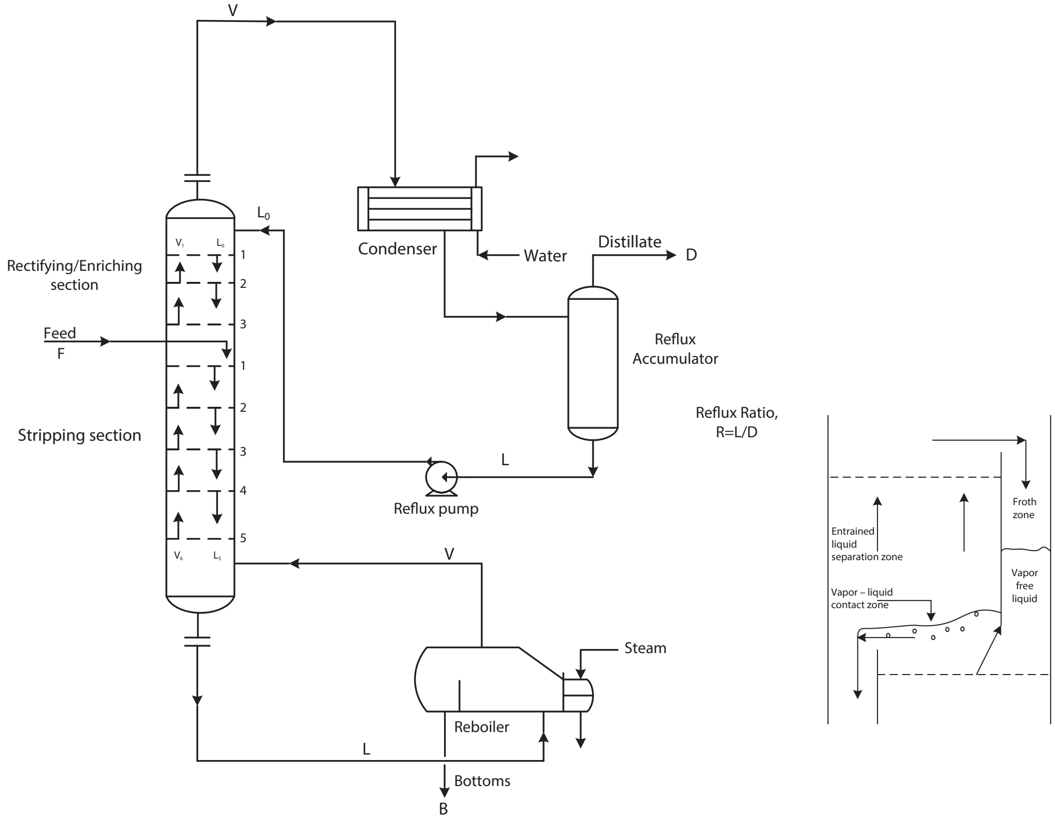


Figure 19.136 Fractionation column with a reboiler and refluxing auxiliaries.

the vapor velocity is raised, froth height increases and approaches the tray above, resulting to froth entrainment flooding.

Downcomer Backup Flooding

Liquid is backed up in the downcomer due to pressure drop in the tray, liquid height on the tray and frictional losses in the downcomer. When the backup in the downcomer exceeds the tray spacing, the downcomer liquid finds a way to the tray above, which causes downcomer backup flooding.

Downcomer Choke Flooding

This is caused if the frictional losses in the downcomer, especially at the downcomer entrance, are high due to high liquid flow rates at the inlet. The liquid accumulates in the tray instead of flowing down, which results in downcomer choke flooding. At the inlet of the downcomer, the volumetric flow of liquid is always higher than at the outlet, as the liquid is more aerated and has a lower density than the liquid at the outlet and consequently requires a larger flow area.

Downcomer Filling: This condition is marked by the downcomer either completely filling or filling to the point that adequate vapor disengagement cannot occur. It can be caused by flooding as indicated earlier or excessive liquid rates. It can also be caused by tray or downcomer restrictions. In a similar fashion to flooding, it can propagate up a fractionating tower.

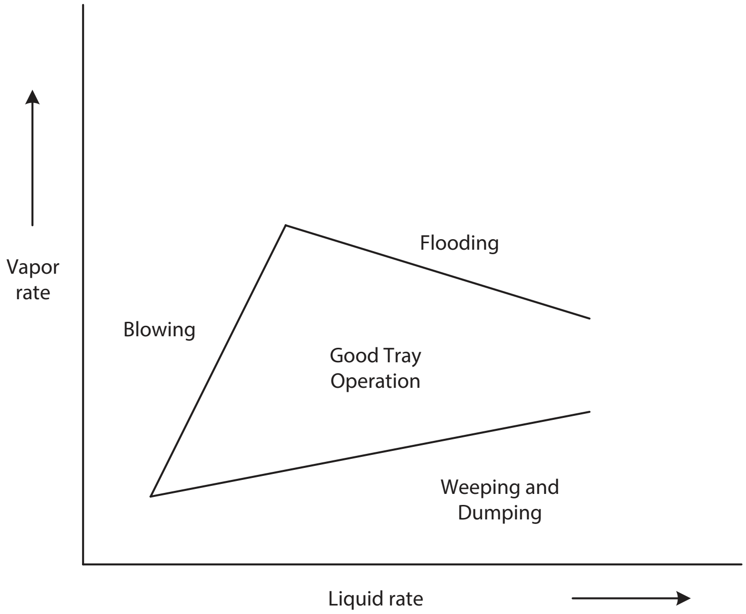


Figure 19.136 (a) A typical tray stability diagram.

Blowing: In this condition, vapor velocity through the holes in a fractionating tray is so high that the vapor phase becomes the continuous phase in the high intensity vapor-liquid contact zone. It is usually caused by a combination of low liquid rates and high vapor rates.

Weeping/Dumping: In this condition, the vapor rates are so low that liquid pours down the holes in the tray. This results in a very low liquid level on the trays, and since the downcomer is no longer sealed, vapor will flow up the downcomer. This single unsealed downcomer will likely cause a high degree of frothing in the inlet downcomer, which will result in liquid holdup in the downcomer. Furthermore, this results in the downcomer flooding in the tray that is weeping/dumping. It should be noted that tray weeping/dumping could also be caused by mechanical damage such as a dropped tray segment.

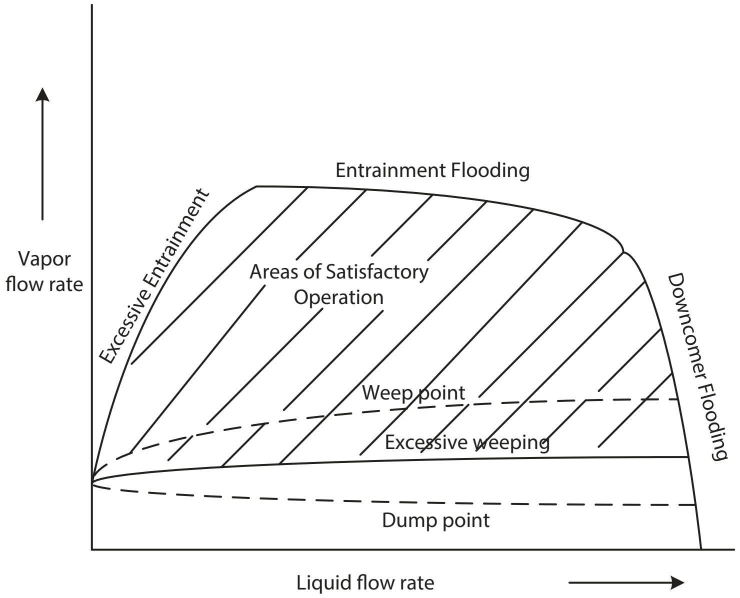


Figure 19.136 (b) Adverse vapor/liquid flow conditions.

Adverse vapor/liquid flow conditions in a fractionating/distillation column can cause the following as shown in Figure 19.136B.

- Foaming
- Entrainment
- Flooding
- Weeping/Dumping
- Downcomer flooding.

Foaming

This refers to the expansion of liquid due to passage of vapor or gas, caused by high vapor flow rates. Although it provides high interfacial liquid-vapor contact, excessive foaming often leads to liquid build up on trays. In some cases, foaming may be so bad that the foam mixes with the liquid on the tray above, resulting in a reduced separation efficiency.

Flooding

Flooding is brought about by excessive vapor flow, causing liquid to be entrained in the vapor up the column. The increased pressure from excessive vapor also backs up the liquid in the downcomer, causing an increase in liquid hold up on the plate above. Depending on the degree of the flooding, the maximum capacity of the column may be severely reduced. Flooding is detected by a sharp increase in column differential pressure and significant decrease in separation efficiency.

Entrainment

This is caused by excessive high vapor flow rates. Entrainment refers to the liquid carried by vapor to the tray above. It is detrimental because the tray efficiency is reduced as the lower volatile material is carried to the plate holding liquid of high volatility. Excessive entrainment can lead to flood.

Weeping/Dumping

This is caused by excessive low vapor flow. The pressure exerted by the vapor is insufficient to hold up the liquid on the tray. Therefore, the liquid starts to leak through the perforations. Excessive weeping will lead to dumping as the liquid on all trays will crash (dump) through to the base of the column and the column would require restarting. Weeping is indicated by a sharp pressure drop in the column and a reduced separation efficiency.

Adverse vapor/liquid flow conditions can cause: foaming, entrainment, flooding, weeping, and downcomer flooding.

Flooding: This condition is marked by vapor velocities being too high that the liquid carried up into the continuous vapor regime does not disengage from the vapor and is carried up to the tray above.

Flooding symptoms:

- An increase of reboiler duty and reflux rate resulting in a decrease in fractionation.
- ΔP above “normal” (Must be reliable ΔP cell)
- For valve trays, “normal” is usually 5–12 mbar per tray, >12 mbar per tray will result in flooding. For packing, 3 kPa/m of packing is high.
- Exception: Tray flooding (entrainment) due to fouling, etc., shows no significant ΔP rise but overhead temperature swing.
- Swings in bottoms rate or level (including side strippers in main fractionator system.
- Liquid is accumulating in tower (e.g., a side draw is reduced and the draw below is not increasing).
- Swing in overhead temperature, often an increase due to liquid entrainment.
- Low temperature profile (i.e., temperature should increase from top to bottom).

What to do?

- Measure the pressure drop (ΔP) over total column. If the column is flooded, then ΔP increases.
- Measure the temperature profile.
- Increase tower pressure to reduce vapor volume (Jet flooding).
- Effective for low-pressure systems.
- Limited by reboiler (temperature must rise).
- Downside: separation loss (lower relative volatility).
- Shift tower heat balance.
- Remove more heat in lower pump-arounds.
- Shift heat from reboiler to feed preheat.
- Probably some separation loss (low reflux rate).
- Reduce stripping steam to the column.
- Increase the column pressure to reduce the vapor loading in the column.
- Reduce the intake to the unit.
- Reduce or cut slops injection.

Entrainment: Caused by excessive high vapor flow rates. It refers to liquid carried by vapor to tray above resulting in a reduced tray efficiency. Excessive entrainment can lead to flooding.

What to do?

- Reduce stripping steam to the column as this will reduce the vapor load.
- Increase the column pressure to reduce the vapor loading in the column.
- Reduce the intake to the unit.
- Scan the column to investigate for low liquid levels.
- Reduce or cut slops injection.

Weeping/Dumping: Caused by excessively low vapor flow that the liquid pours down the holes in the tray. This may result in downcomer flooding in the trays above. Excessive weeping may lead to dumping the liquid on all trays, which will crash (dump).

What to do?

- Increase stripping steam to the column.
- Decrease overhead condenser capacity (reduces the liquid quantity).

Figure 9.137 shows schemes of normal operation, liquid and jet flooding and weeping in a column.

Fractionation Problem Solving Considerations

In assuming that the fractionating tower is being designed correctly, problem solving will always involve process changes or mechanical damage. Confirmation that the fractionating tower was designed correctly can be ascertained by developing a tray stability diagram as discussed earlier. Some possible mechanical damages that could be considered are as follows:

- Failure of a tray segment
- Uneven trays
- Manways blown
- Crushed packing
- Restrictions
- Vapor inlet changes

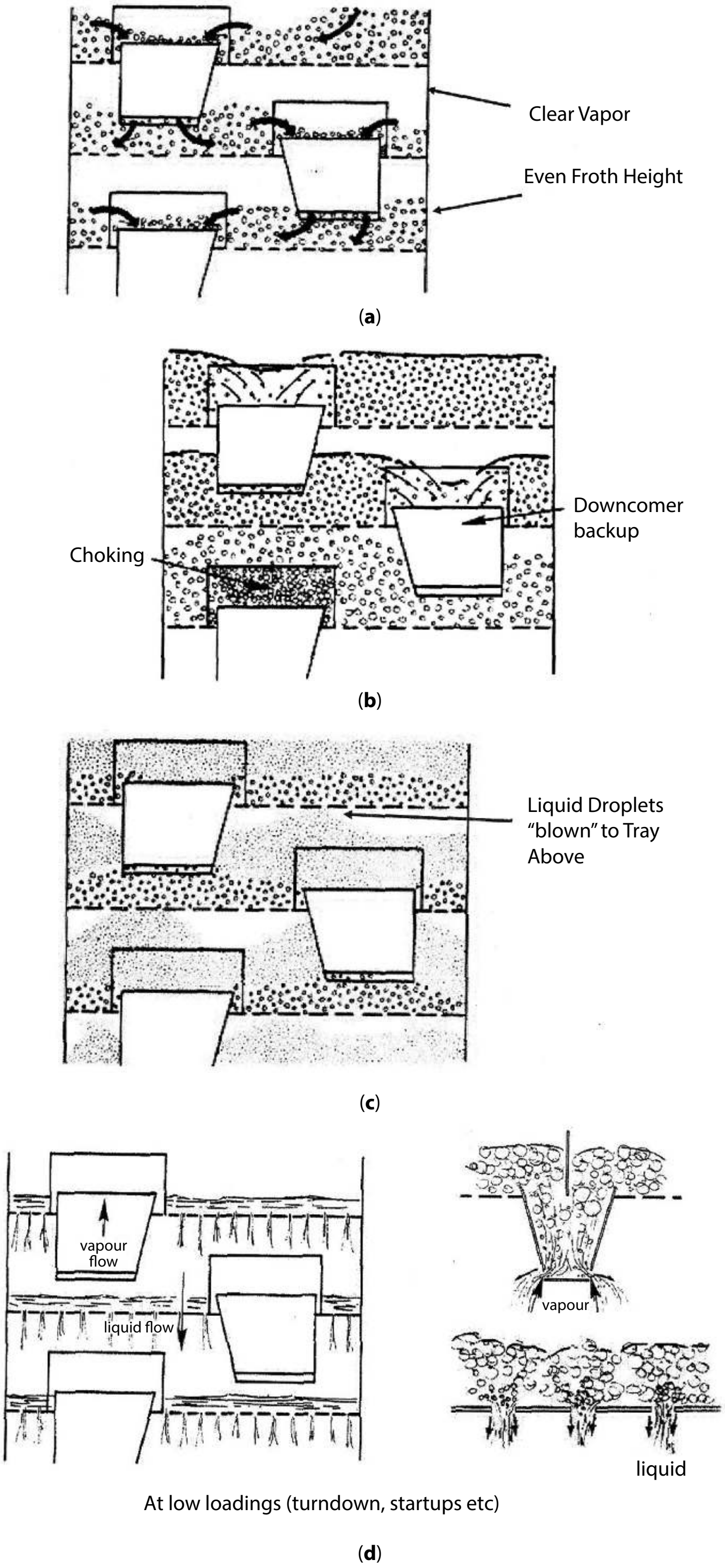


Figure 19.137 Schemes of (a) normal operation, (b) liquid flooding (choking), and (c) jet flooding (entrainment), and (d) weeping and unsealing in the column.

Process changes that cause poor tray performance may be due to known changes such as an increase in rates or very subtle changes such as follows:

- Instrument errors causing excessive vapor or liquid rates.
- Foaming caused by trace quantities of a surface-active ingredient.
- Plugging caused by the presence of water which reacts with a soluble material to form solids.
- Control coupling.

Process engineers should consult with panel operators and seek their opinions as to what has changed since the time that the column was operating satisfactorily. These involve the following:

1. Verify the suspect instrumentation.
2. Check the pressure and temperature gauges (Figures 19.138).
3. Measure the column section pressure drops.
4. Check the product qualities.
5. Check the operations against the design and last test runs data.
6. Obtain a clear and factual definitions of the symptoms.
7. Analyze the gathered field data (e.g., ΔP data).
8. Study the column response to changes in the level, reflux rate and reboiler duty.
9. Identify possible causes and corresponding tests, which will confirm or refute them.
10. Scan the column if required (Figures 19.139a–e).
11. Involve distillation specialists.

19.53.1 Common Problems in Distillation Columns [225]

Problems are encountered in distillation, which are not usually caused by the actual theories in the design but by the difficulties or errors in its internals. The following are the most common problems [225]:

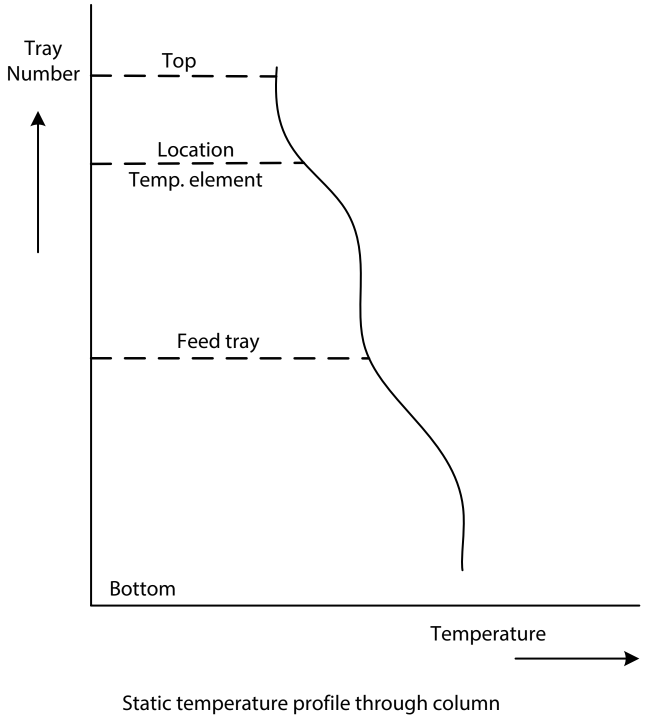


Figure 19.138 Temperature profile vs. number of trays.

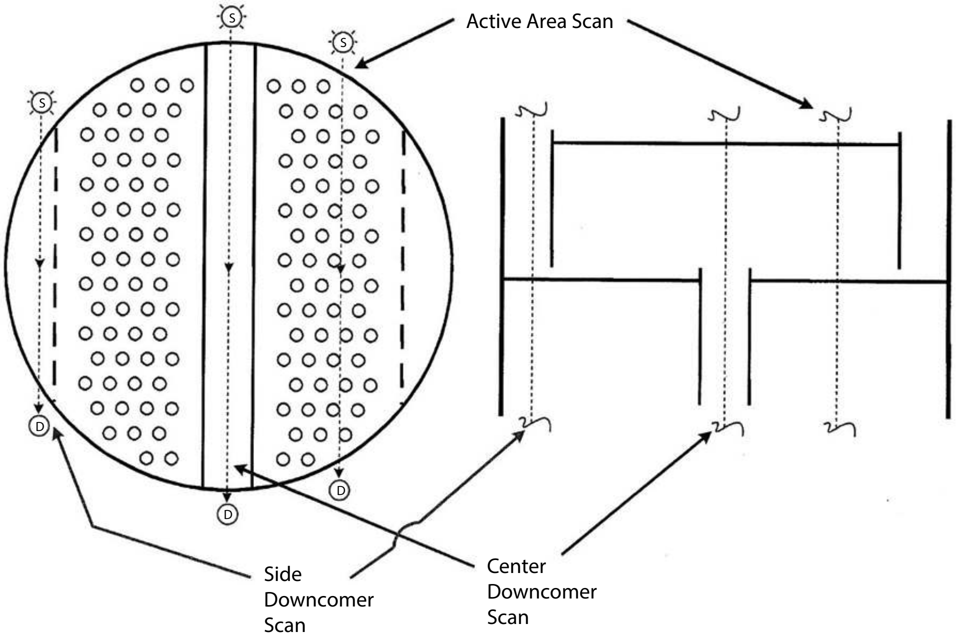


Figure 19.139 (a) Potential scan lines on a two-pass tray.

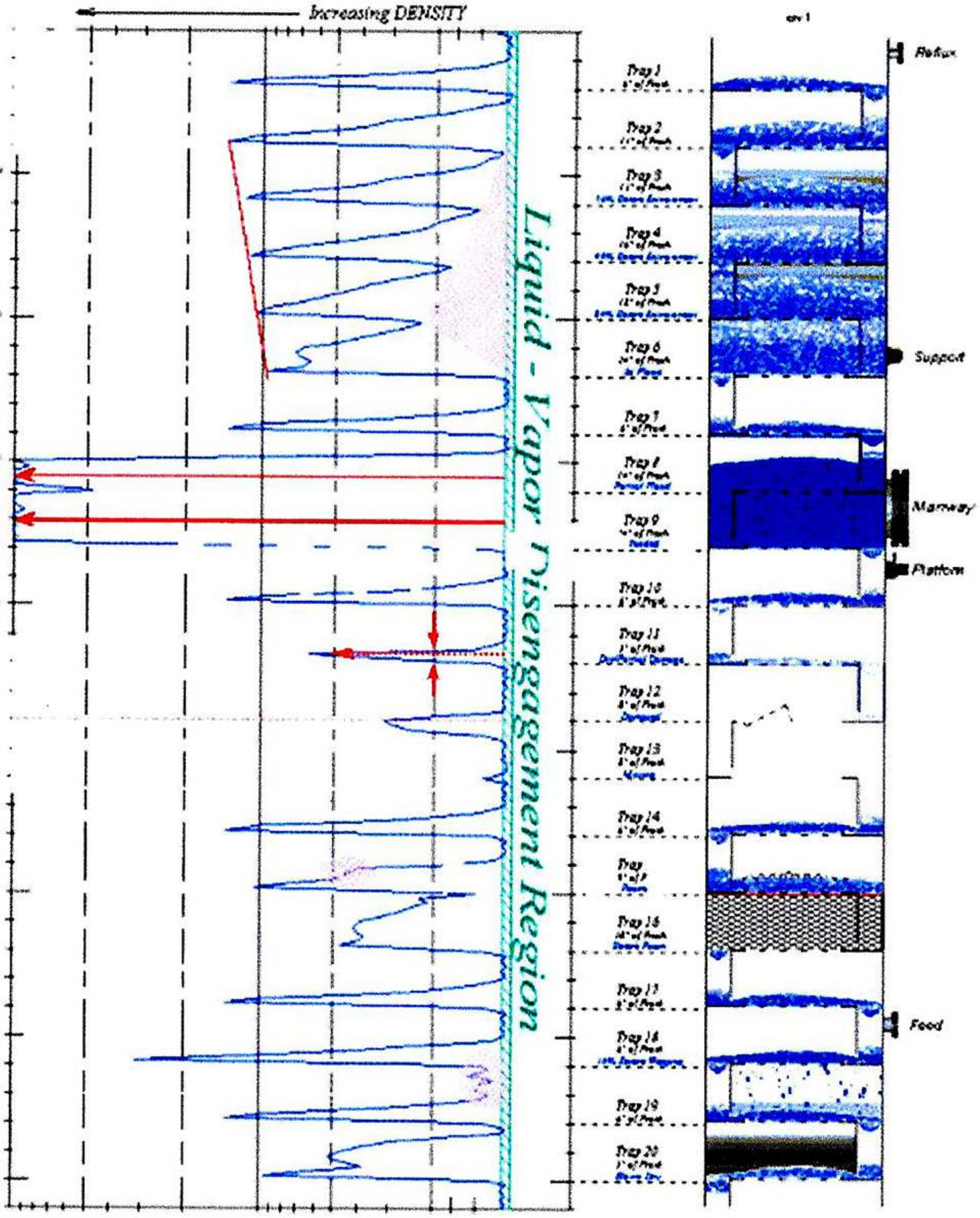
1. Incorrect fitting of the tray decks with tray support rings (TSRs) and support beams, which results in heavy weeping.
2. Light overloading on trays in (gpm/in. of weir), which results in downcomer choke flooding.
3. Inadequate downcomer bottom clearance, which leads to high backup and flooding.
4. Sticking or clogging of tray dispersers (like valves, etc.), which causes a high pressure drop in the column or section of the column.
5. Heavy foaming, which results in poor separation.
6. Often short downcomer residence times result in the recirculation of vapor (unreleased) from the tray above to the tray below along with the downcomer liquid. This results in severe loss of efficiency due to back mixing.
7. Water in reflux leads to pressure fluctuations in a column and results in poor separation. The most common cause of water ingress in a distillation column is improper separation of water in the reflux drum. Sometimes a small leak in the water condenser can result in more water entering the reflux drum (for low-pressure columns) and can get into the column along with the reflux. An incorrect measurement of the water–hydrocarbon interface level can also lead to water in the reflux.

   Steps to avoid water in the reflux:

   a. Reflux draw pipe should always be extended 6 in. inside the reflux drum.
   b. Adequately size the reflux drum for water–oil separation (see Chapter 18).
   c. Adequately size the water boot.
8. Dislodgement of a tray deck or sections can be caused by the following:
   a. Condensate in stripping steam: Tray decks/segments often get dislodged due to an improper original fitting, water shots, or pressure surges during operation and startups. Chronic problems have arisen with the dislodgement of the atmospheric and vacuum columns' stripping section trays. The stripping agent is superheated steam and four to nine trays are provided in the bottom stripping section. The trays can become dislodged during startup when the stripping agent is introduced in the column, as the stripping steam may contain condensate due to a stagnant condition of the lines during shutdown. If this is the case, the stripping agent will put water in the hot column, inducing a surge and causing consequent tray damage.

b. High liquid level in column bottom: If the liquid level becomes high in the bottom of the column and goes above the reboiler vapor returns or stripping steam inlet nozzle, the vapor may travel through the liquid pool in the form of slugs. This may damage the bottom trays and tray supports and can result in tray dislodgment.

This can be resolved by stopping the reboiler return vapor/stripping steam flow in case of high levels in the bottom sump by way of terminating reboiler hot fluid or shutting off the stripping steam valve. Another reason for dislodgment of column bottom tray is quick pump out of the built-up level, a phenomenon often referred to by experts as a vapor gap. The liquid from upper trays cannot drain out as fast as it is pumped out of the sump. A vapor gap is generated below the bottom tray, and when the load of accumulated liquid comes on the tray, it fails to support and



Proprietary and Confidential – Gamma Surveys – Copyright 1999 - 2012 – All Rights Reserved

Figure 19.139 (b) Gamma scanning of a distillation column.

dislodges [204]. Cross-flow trays are susceptible to this kind of failure where downcomer clearance and downcomer flow area are small.

Often bottom trays (particularly those in crude and vacuum columns) are changed to counterflow trays without downcomers to alleviate the problem. It may be better to use counterflow trays (baffle trays, disc and doughnut trays), as there is a bigger flow area for the accumulated liquid to flow down and reduce the likelihood of vapor gap and tray dislodgment. However, repeated failures of counterflow trays in the column bottom are still observed. It is likely that failure can be avoided by strengthening the trays and lowering the rate of pump-out.

9. Absence of self-venting nozzles and lines in product and pump-around draw-offs: Draw-off nozzles and lines must meet self-venting criteria. The liquid drawn from the column trays still has some amount of gas (aerated liquid) dissolved in it. If the same liquid is fed to pumps, they are likely to

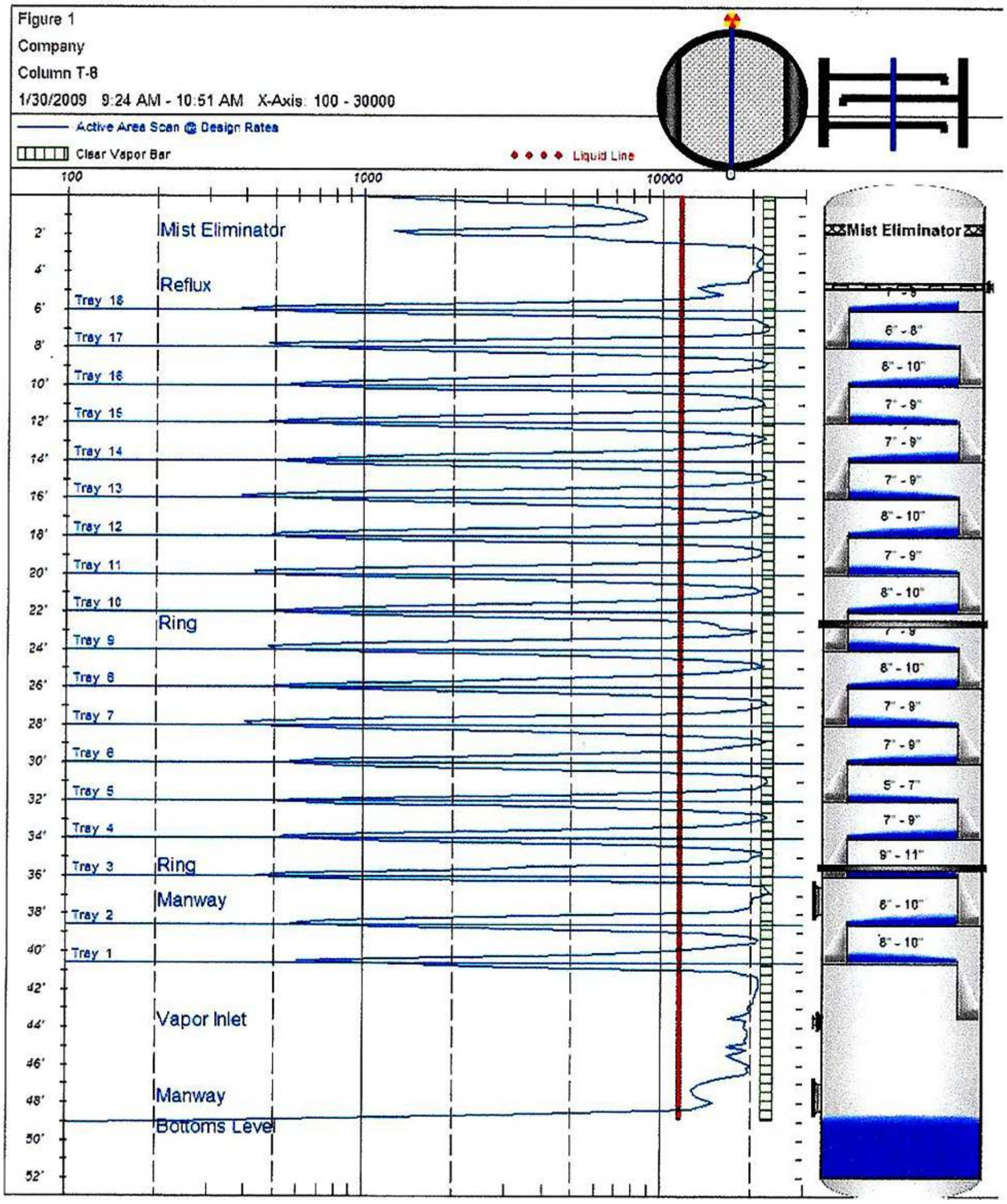


Figure 19.139 (c) A typical gamma scanning plot of a column.

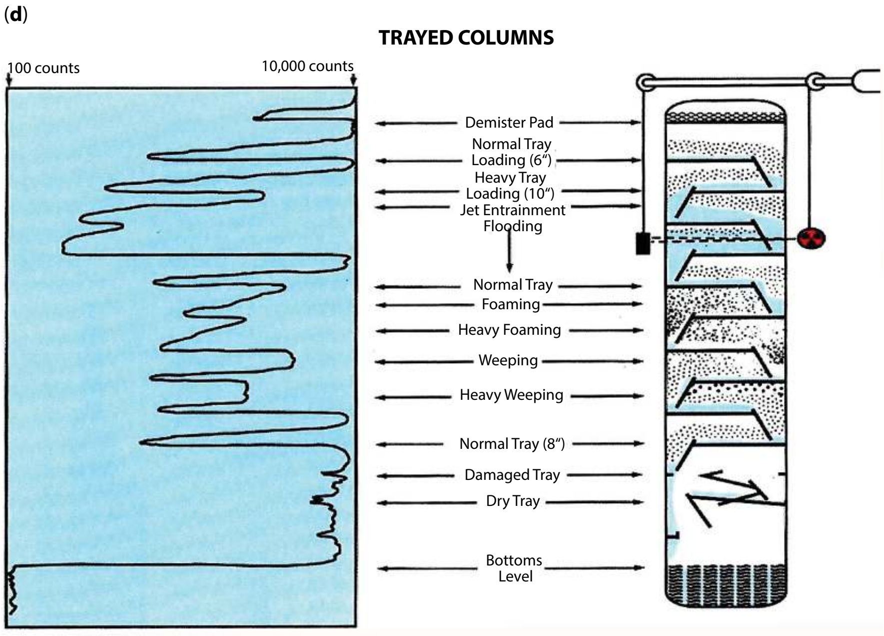


Figure 19.139 (d) Gamma scanning of a trayed column.

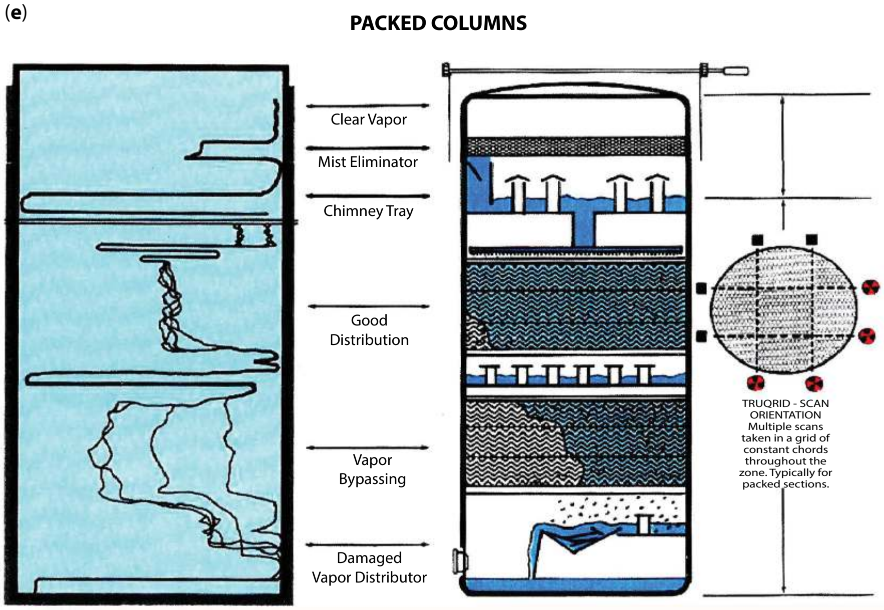


Figure 19.139 (e) Gamma scanning of a packed column.

cavitate with the release of gas from the liquid at pump suction (see Chapter 15). The nozzle and connecting line size must be adequate to allow the release of the gas back into the column. After around 10 ft. of vertical travel, the withdrawal line can be swaged in order to reduce the diameter. This will help release the vapor back into the column.

10. Incorrect construction of draw-off channels/draw-off sumps: This is a very common problem encountered in distillation columns in refineries, such as in crude and vacuum distillation units where channels and draw-off sumps were too small to handle the liquid load. This resulted in the unavailability of pump-around reflux flows. The problems were subsequently corrected as per the guidelines below.

Guidelines for designing a draw-off box:

- **a.** The depth of the draw-off box should be 150 mm or 1.5 times the diameter of the withdrawal nozzle, whichever is greater.
- **b.** The draw-off box area should allow a maximum downward velocity of 1 ft/s.
- **c.** The velocity in the channels should be at maximum 1 ft/s., as non-compliance can result in unavailability of flow for pump-around refluxes in refinery column.
- **d.** It is also very important that the longer sides of the chimney be parallel to the liquid flow direction toward the withdrawal box or collector channel (which eventually lead to the draw-off box); otherwise, the flow of liquid may be restricted.

11. Inadequate residence time at column bottom/draw-offs: Residence time at column bottom sumps and draw-off trays is a very important parameter, especially when the bottom liquid is fed to a furnace. Loss of suction of the pump due to unreleased gas may coke up the furnace. If residence time in the spout is low, this problem can be resolved by external vessel breathing with the column vapor space above the liquid.

Another option is to provide a liquid gallery so that liquid from the bottom most tray can flow through if (while releasing vapors) and reach the bottom. People often use a self-venting nozzle and line for the bottom products going to pump in the absence of an adequate bottom residence time. However, this may not be a foolproof arrangement.

When sufficient residence time is possible, the nozzle size can be smaller than the self-venting size. In revamps, such nozzles are commonly encountered. In one naphtha splitter producing ISOM feed and CCRU feed with a furnace reboiler, the nozzle size was allowed to remain the same after it was established that the residence time at the bottom was adequate.

12. Hydrates in trays: Hydrates are likely to form in natural gas processing plants in the distillation section. The hydrates can choke the tray dispersers. This increases the pressure drop and leads to column flooding due to the partial plugging of the dispersers.

13. Low clearance between tray deck and downcomer/downcomer edge with inlet weir during fitting: Often tray inlet weirs are used by internal experts to ensure a better distribution of liquids, particularly where liquid flow rates are low. The horizontal clearance between the downcomer and the inlet weir needs to be maintained such that the downcomer backups is within allowable limits of 40–50% of tray spacing.

14. Coking in trays and choking due to the deposition of salts: Coking can occur in trays when hydrocarbons degrade, polymerize, and form deposits. Often components present can combine to produce salts and deposits on trays. Sometimes particulates carried from upstream deposit on tray dispersers, reducing the flow area for the vapor. The liquid channels may also become blocked due to the deposits.

Figure 19.140 shows the flow chart of problems relating to a heavy gas oil from the column.

The following are troubleshooting checklists in the distillation column (Figure 19.141):

- Do we have a good heat and mass balance?
- Have we checked the key instruments?
- All are levels in the tower normal—how can we confirm this?
- Is the operating pressure normal—what is the column differential pressure?
- Is the tower ΔT normal—what does this tell us?
- Is there any variation in feed quality and composition—are feed analyses carried out?
- Is the system foaming—jar test?
- Is there water in the unit or in the feed—how can we test for this?
- Could we have a build-up of inerts or impurities in the column—would this persist over time?
- Extraneous feeds into the column—leaking heat exchangers?
- Any indication of mechanical damage to internals—tray parts in the suction screens? Do we require a gamma scan?

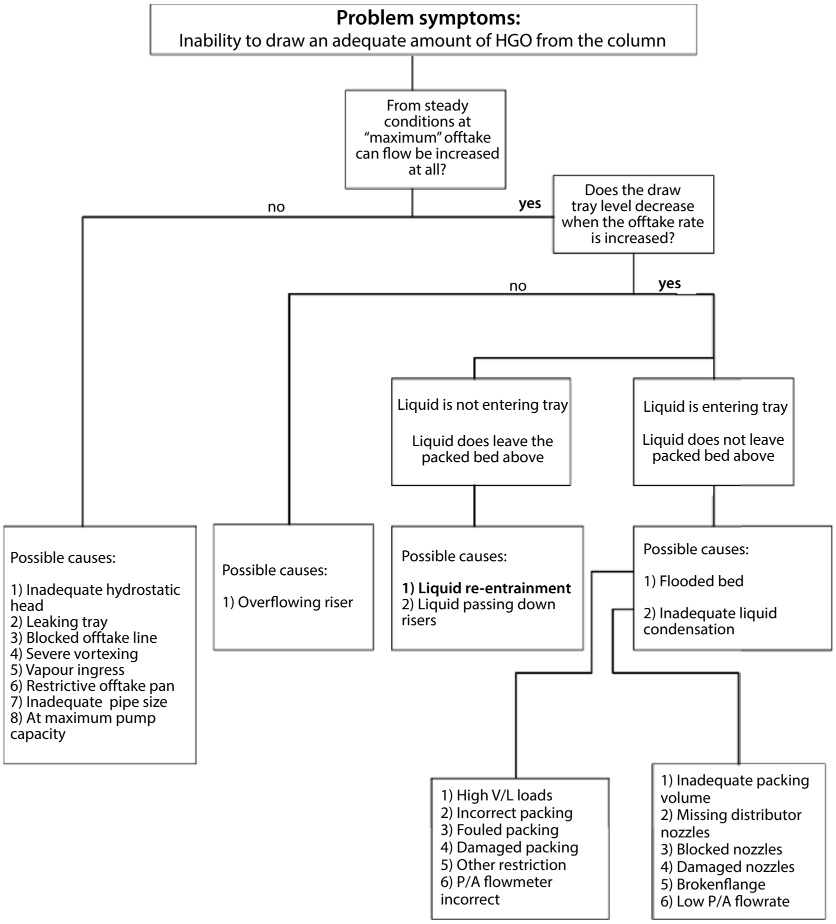


Figure 19.140 Problem symptoms relating to heavy gas oil from the distillation column.

- Any indication of plugging or fouling of the internals, exchanger ΔP?
- Any problems associated with utilities?

Here are the troubleshooting tools in determining problems in the distillation column.

- Make use of the plant history and review process data vs. time.
- Understand the operating and technical fundamentals of the column.
- Know how a column functions, hydraulic constraints, thermodynamic, and equilibrium limits.
- Employ a single pressure gauge survey or differential pressure (DP) measurement on the column to confirm whether flooding occurs. Can usually find some spare nozzles to install DP transmitter. If not, use a calibrated single gauge with the appropriate span.
- Apply pressure or DP indicator to confirm the operating level in the column.
- Acquire the heat and material balances.
- Withdraw samples and use analytical data to help isolate the problems.
- Use temperature indicator/profiles in ascertaining the column is performing correctly.
- Use a non-intrusive radiation technique.

Fundamentally, there are at least three types of distillation problems, namely:

1. Inappropriate design.
2. Inappropriate operation.
3. Potential damage to internal equipment.

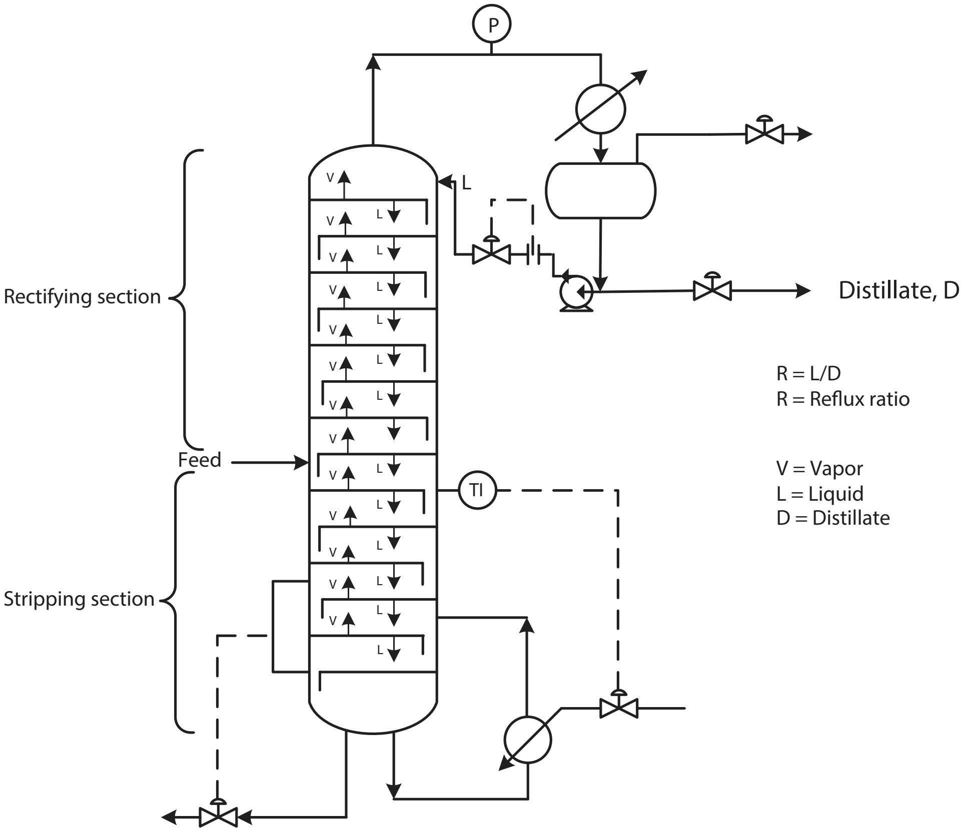


Figure 19.141 A trayed distillation column with auxiliary equipment, instruments, and fittings.

Before a process is shut down for maintenance/repairs, the inappropriate design and damage to internal equipment should be reviewed and determined and inappropriate operation should be eliminated.

Column Profiles

Composition Profile: More light material moves higher up in the column.
Temperature Profile: The temperature increases down the column (Figure 19.138).

Composition and temperature profiles are essential as potential problems in the column are:

- Flooding
- Column internals damage/Installation problem (Figures 19.142 and 19.143)
- Maldistribution.

Pressure Profile

The following parameters are carried out during troubleshooting of the distillation/fractionating column:

- The hydraulics of a distillation column:
  - Liquid flows down the column by gravity.
  - As the liquid descends it takes flowing ΔP losses through the column internals.
  - Vapor flows up the column seeking the low pressure "sink"

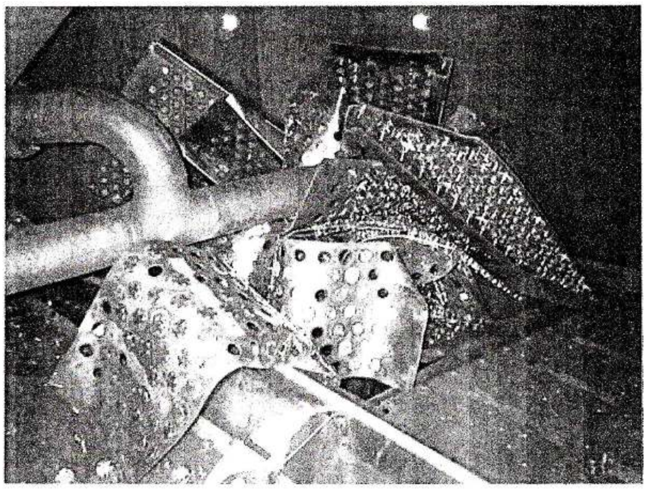


Figure 19.142 Column internal damage.

Figure 19.143 Gasoline splitter -installation problem of the column. Column suddenly lost all separation.

  - (condenser/accumulator or overhead vapor destination)
  - As the vapor rises, it takes flowing ΔP losses through the column internals and bubbling through the aerated liquid levels on the trays.

- The resulting pressure profile is important because:
  - It helps to indicate how loaded the internals in the column.
  - It locates the problem areas in the column.
  - ΔP shows how much liquid is being held up on the internals.

Typical Pressure Drops (Approximate)

Packing

Random, < 1 in. H_2O per vertical ft (< 83 mm H_2O per m).
Structured, < 0.3 in. H_2O per vertical ft (or about 1 psi for 100 vertical ft. of packing) (< 25 mm H_2O per m).

Trays

– 0.15 psi (3-4 in. H_2O or 75 – 100 mm H_2O) per tray total caused by:
~ 1/3 due to "dry" pressure drop through the valves or sieve holes
~ 2/3 due to bubbling through the froth.
The actual froth is aerated, and is typically from 4–12 in. (100–300 mm) deep on the tray.
A typical 50 trays column would have ΔP of 5–8 psi (0.34–0.55 bar)
Condenser typically has ΔP of 3-10 psi (0.2–0.7 bar)

Figures 19.144 and 19.145 show profiles of liquid and vapor flow rate vs. theoretical stage number and crude tower tray composition vs. theoretical number of stages respectively. Figure 19.146 shows tray fouling in the column resulting in reduced separation and efficiency.

Appropriate Stage Design

Computer simulation packages are readily employed in the design of distillation/fractionation columns as compared to the trial-and-error basis. The computer model is expected to match the existing field data if the column

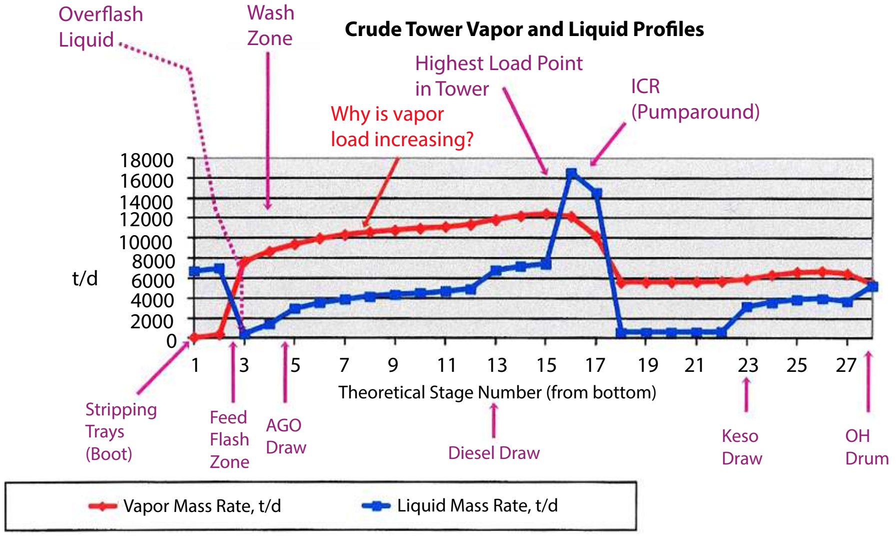


Figure 19.144 Profiles of liquid and vapor flow rate vs. theoretical stage number.

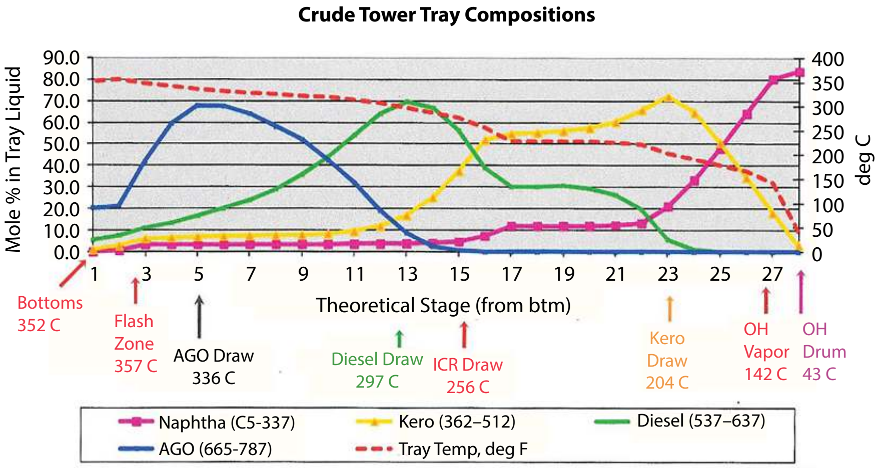


Figure 19.145 Crude tower tray composition vs. theoretical number of stages.

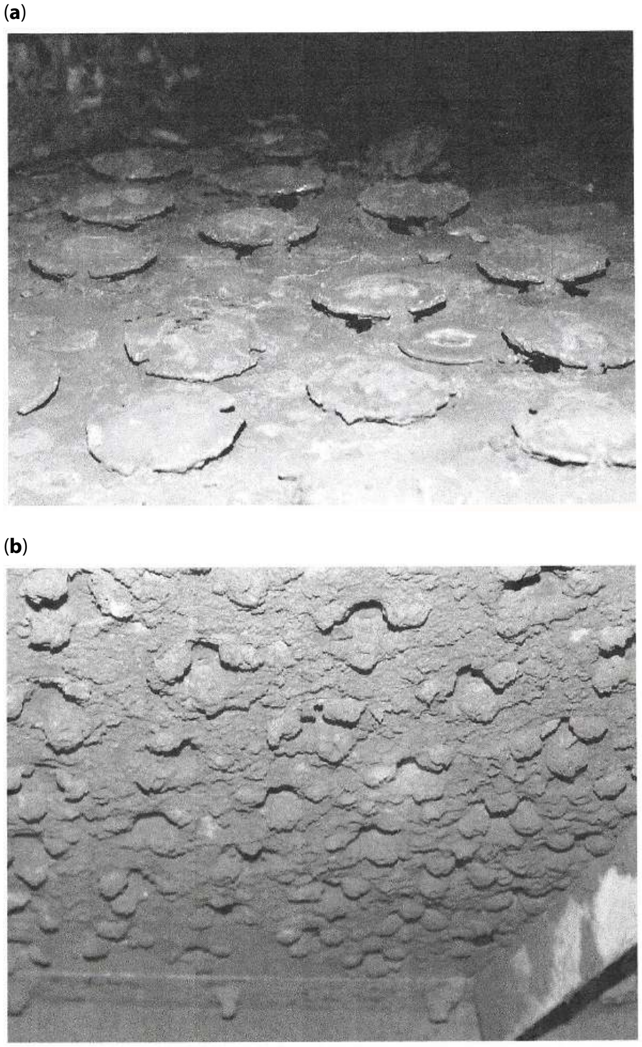


Figure 19.146 (a) and (b) Tray fouling.

is operating as designed, otherwise the field data should be reviewed and verified. If the field data are accurate, the model should be adjusted to match the existing data. The computer model can be verified by developing a McCabe-Thiele plot to verify the number of separation stages required. If we assume that the model is correct in all cases, this could have a deleterious effect in the operation and performance of the column.

Trayed columns utilize a pressure and temperature differential to create a mass transfer gradient to separate the products. Packed columns generate a mass transfer area by providing a large surface area over which the liquid can transfer heat and mass to the vapor.

Pressure is an important parameter and thus acts as a constraint in the stage design. In low pressure systems, the vapor is considered as the continuous phase while in high pressure systems, the liquid is considered as the continuous phase. In low pressure systems, packing is successfully utilized and in high pressure systems, this can fail due to a back mixing mechanism. Therefore, trays are preferred.

The temperature at which the reflux can be condensed usually determines the column pressure. Normally, the preferred temperature is that of cooling water. If cooling water cannot be used, the pressure is set using the best combination of variables. The first variable is that hydrocarbons separate easier at lower pressures due to the difference in their relative volatilities are larger at lower pressures. Thus, a lower pressure equates to less reflux, less stages and smaller columns.

The second variable is the increased cost of cooling at lower pressures. At this condition, refrigeration is required resulting in an increased operating cost. A balance has to be constructed between capital cost and operating cost. This balance is used in the determination of the amount of reflux against the number of stages. This balance is between the energy requirement versus the cost of the additional stages. The use of computer simulation has enabled this balance to be optimal.

Pressure Choices

During the design of a distillation column, the controlling factor in choosing the pressure of the column is the heating requirements of the reboiler or the condensing requirements of the overhead condenser. For example, a new column is installed to separate benzene and toluene. Here, the controlling factor in choosing the column pressure is the ability to use low-pressure boiler feed water as a condensation medium to produce low pressure steam. This choice would mean that the column is going to operate under pressure. Performing this separation under pressure has the following advantages (366):

- Increasing the column pressure would increase the vapor density and thus the vapor handling capacity. This results in a smaller column diameter, thus lowering the overall cost of the project.
- It would allow the possibility of having the benzene-toluene splitter share a condenser with a column used to remove benzene from the vent gas. Both of the column overheads products would flow to the same location. The cost of installing a complete condenser system for this column would be reduced.

Increasing the column's operating pressure would present some unfavorable effects such as follows:

1. Raising the pressure lowers the relative volatility, makes the separation more difficult.
2. Raising the temperature also raises the reboiler temperature, thereby requiring a more expensive heating medium. A reboiler with a larger heat transfer area would be required.
3. A pressure > 100 psig, requires a corresponding increase in the column's shell thickness, which invariably increases the capital cost.

Reboiler selection and design are described in chapter 21.

Rating of Columns

The chemical/process engineer determines a conservative rating of an existing column by providing the actual diameter, spacing and expected system factor, and then calculates the flood percentage. Commercial simulators have the capacity to rate the trays by inputting the column and hole diameter, the tray spacing, the number of liquid passes, and so on. The percent flood, pressure drop, and downcomer backup are also calculated.

If the percent flooding is $\leq$ 70% for a column with a diameter of less than 2 m, then the column diameter is considered satisfactory. The downcomer backup is also further checked, and must be at a maximum 40% to 50 %. For columns with larger diameters, a flood percentage of up to 80% is considered acceptable for the preliminary rating of the column. The downcomer backup is also further checked and must be at a maximum 40%-50%. The final rating of the internals is carried out by column design experts. The chemical/process engineer should have some understanding of the fundamental principles of column diameter estimation as this ensures that by performing this rating, it enables him/her to troubleshoot and perform preliminary checks.

Column diameter is estimated based on the type of internals, the system factor, the tray spacing (normally 600 mm) and the percentage of flooding specified.

1. 70 % for columns with a diameter of up to 2 m.
2. 80% for columns with a diameter greater than 2 m.

The estimation of column diameters needs to be conservative, however, the design experts will match the pressure drop and the stages specified during detail design of the internals.

Troubleshooting of Distillation Columns with Trays [331]

The trayed columns utilize a pressure and temperature differential to separate the products. The weir holds a liquid level of each tray as the vapor must overcome this liquid head to move up the column. On each tray, the vapor rising up the column makes contact with liquid flowing down the column, and then above the tray, they are separated. Any deviation that develops by restricting the vapor and liquid from contacting and separating will prevent the column from achieving its required design specifications.

Deviations that will restrict the ability of the vapor-liquid contact in the column include the following:

1. Corroded, fouled, or eroded tray valves and feed devices.
2. Restrictions in downcomers.
3. Physical damage.

Deviations that will restrict the ability of the column to separate include the following:

1. Excessive rates.
2. Contaminates that cause foaming.
3. Improperly sized downcomer openings.
4. Liquid entrainment.
5. Foaming
6. Excessive liquid back-up in the downcomers.
7. Physical damage.

Because the trayed column uses a pressure and temperature differentials to separate the products, the pressure and temperature profiles of a column are a key indicator of how the column is performing. The column temperature and pressure should gradually increase along the downward length of the column.

A calculation can be developed for what the column ΔP should be based on the number of stages and the height of the weir, or the downcomer dam. For example, if a column has 20 trays and a weir height of 3 in., the vapor has to overcome a liquid height of 60 in. the equivalent height of one pound of H_2O is 27.68 in. Therefore, 60 in. of H_2O equates to a ΔP of 2.16 psi. This calculation is corrected by the specific gravity of the actual liquid on the tray.

The ΔP calculation is an essential tool in tower troubleshooting. If ΔP is low, tower may be weeping, internal tray man-ways may be dislodged or the reflux flow may be inaccurate on the low side. If ΔP is high, the downcomers may be restricted, the whole trays may be dislodged, or the reflux flow rate may be inaccurate on the high side.

After the calculation of the ΔP, a pressure survey should be carried out using the same pre-calibrated gauge. Gauges can be damaged in installation, and due care should be exercised in the installation of the gauge. Field mounted pressure gauges contain a high level of uncertainty and should not be used for troubleshooting.

Pressure Survey

Pressure drop ΔP is proportional to V^2, Q^2 and $(1/A^2)$.

where
V = linear velocity, Q = volumetric flow rate, A = cross-sectional area for flow.

Temperature Survey

A temperature profile will also provide valuable information as to the operations of the column. If thermo-wells and thermocouples are not available at the desired points throughout the tower, then IR temperature scan gun can be used at the inspection ports through the insulation. The temperature profile is surveyed around the column at:

1. Feed
2. Bottoms
3. Overhead temperature
4. Reflux temperature
5. Reboiler heat input
6. Overhead receiver cooling in and out.

If the feed is hotter than the design, then it could form vapor and resulting in flooding the upper part (i.e., the rectifying section) of the column. If the feed is colder than the design, then it can form liquid and flood the lower part (i.e., the stripping section) of the column. If the reflux is much hotter than the design, then the liquid is recirculating around the top part of the column instead of traversing down the column (see Figure 19.141).

If the reflux is much colder than the design, then the top part of the column can be flooded.

If the cooling water temperature is the same as in and out of the condenser, then it shows that there is insufficient overhead vapor from the top of the column.

If the temperature profile is not consistent, then several causes are possible, namely:

1. Hydraulic tray flooding or weeping.
2. Potential tray damage.
3. Fouled or corroded trays or downcomers.

In distillation towers, there are two accumulators. The first is the overhead receiver and the second is the bottom section of the tower. These accumulators are used to stabilize the operation of the tower and downstream operations. This internal surge drum creates an inventory to act as a buffer. If the internal level is allowed to rise above the reboiler return stripping inlet or feed inlet, then flooding can occur.

There is an inherent error built into sight glass and level instrumentation. The sight glass and level instrumentation contain non aerated liquid called clear liquid, which is not a true indication of the condition of the liquid within the column. The liquid within the column will have two levels, a clear liquid level below the aerated liquid level. Because the aerated level will have lower specific gravity than the clear liquid within the instrumentation, the column level will be higher than the instrumentation. If the level in the tower is higher than the feed or reboiler return entrained liquid can be carried to the next stage causing flooding in the column. This is the principal cause of the BP refinery Texas explosion in 2005, and a case study is reviewed in the text.

Heat Transfer Rate Calculation

$$Q = U_D \cdot A \cdot \Delta T_{LMTD} = M \cdot Cp \Delta T$$

where

A = surface area for heat transfer
ΔT_{LMT} = log mean temperature difference
Cp = specific heat
ΔT = temperature differential
M = the mass flow rate of the fluid

Synchronous Speed Calculation

$$n = (120\ f)/P$$

where f = frequency in cycles/sec., n = synchronous speed (rpm) and P = number of poles in the motor.

Use of Affinity Laws (Centrifugal Device)

Flow rate Q proportional to n, d
Head (H) proportional to n^2, d^2
Power proportional to n^3, d^3
n is speed in RPM and d is the impeller diameter
Net positive suction head (NPSH) available at the suction of a pump = $(P_s/\rho) - (P_v/\rho)$

where

P_s = Pump suction pressure
P_v = Vapor pressure at pumping temperature
ρ = Density at pumping temperature

For liquid at its bubble point, the elevation of the vessel and column containing the liquid only provides the NPSH available as the suction pressure and vapor pressure cancel out.

Synopsis of Tray Troubleshooting

The checklist is as follows:

1. Ensure that the levels are accurate. Have operations move levels and view changes in the field.
2. Calculate the column ΔP, and then measure the ΔP. Review the survey pressure readings to the operation's readings.
3. Survey column temperature profile. Review the survey temperature reading to operation's reading.

Verify the Column's Operations

Perform the tower simulation to verify the tower stage efficiency. Sometimes, the feed compositions change and the tower is no longer able to meet the desired specifications due to thermodynamic or equilibrium constraints. The requirement to perform the simulations will be as follows:

- Accurate tower feed, overhead and bottoms laboratory analyses.
- Accurate tower mass balance, within 2%.
- Heating and cooling medium temperatures.

If the tower simulation confirms the limits are not beyond the thermodynamic or equilibrium constraints then an additional check may be to scan the column for tray damage. This type of troubleshooting method can determine internal damage, vapor-liquid maldistribution, and packed and trayed tower fouling. Because of the economic constraints, scanning should be the last resort after the simple checks and the limits are confirmed. This technique can also confirm the problems that were identified by the other checks.

Good Design Practices

General Guidelines

The following general guidelines in distillation and ancillary equipment are as follows:

1. Velocity of fluids
   - Coke drum vapor space velocity: 0.3–0.6 ft/s
   - Column overhead line velocity: 15–20 m/s
   - Rich amine line velocity: < 1.0 m/s
   - Lean amine line velocity: < 2.0 m/s
2. Column draw-off nozzles and lines are required to meet self-venting criteria. This is required so that the dissolved gases do not get into the draw-off pumps and result in pump cavitation.
3. Residence time can be maintained in the column or vessel bottom spout when the liquid is picked up by a pump and fed downstream. A vortex breaker may also be advisable (see Chapter 15). Adequate residence time is required to eliminate the likelihood of unreleased vapor going to the pumps and causing pump cavitation. The residence time should be estimated based on the volume of liquid between the low–level alarm (LLA) and high level alarm (HLA). Typical residence time ranges from 2 to 10 min.
4. Compressor suction lines are to be made self-draining toward the suction knock-out drums. Suction lines are to be designed in accordance with API 617 or API 618.
5. Equipment downstream of any pump is to have the minimum design pressure equal to the shutoff pressure of the pump + suction equipment maximum pressure (safety setting) + suction height.
6. Highly oversized safety valves may lead to column internal damage due to higher superficial velocities. It may be better to set the design pressure of the equipment and safety valve higher (minimum at 3.5 $kg/cm^2 g$). This will reduce the specific volume of the vapor at safety discharge pressure conditions and would reduce drag force (proportional to ρV^2) on internals. ρ is the vapor density at the pressure and temperature and V is the vapor or gas velocity.

A Case Study

Troubleshoot Flooding Problems in A Crude Distillation Column

Ozturk *et al.* [237] investigated a crude distillation unit (CDU) that produces naphtha, kerosene, light diesel, heavy diesel, and atmospheric residue at Tupras, Izmit refinery. The CDU was operated under optimized, stable conditions to provide on-specification products and high refinery profits. However, expected operating conditions were not achievable due to design limitations of the main equipment, which resulted in the inspection of the main crude distillation column, as emergency shutdowns might pose other problems in the unit.

The column suffered from temperature profile variation after the unit restarted following an emergency shutdown. Furthermore, the variation is column temperature profile led to unstable and inefficient operation, reduced feed rate and off-specification products. The cause led to a step-by-step investigation to find a solution to the problem. Process data showed that flooding was the culprit. Optimum working conditions were provided after necessary maintenance work was carried out.

The CDUs at Izmit refinery produce mixture of LPG, light and heavy naphtha, kerosene, light and heavy diesel, and atmospheric residue. The oldest of the CDU has the lowest capacity of the three. It removes salt and water from the crude oil in the desalter section and then separates the crude into products according to their boiling point differences in the atmospheric distillation section. LPG and naphtha are sent to the naphtha stabilization section and atmospheric residue is processed in the vacuum distillation unit (VDU).

The CDU experienced an emergency shutdown and underwent maintenance. However, upon restarting, it was observed that the temperature profile of the CDU oscillated. This temperature fluctuation limited unit operation, especially during the processing of crude oil with high naphtha. During the turnaround in October 2019, plugged nozzles were observed on the top reflux distributor, and holes were found in the downcomer where the top reflux discharges. A maintenance work which was carried out on the top reflux system was assumed as the reason for the oscillation problem, but it continued after the unit was restarted.

To determine the reason for the temperature fluctuation, the unit was shut down again. Upon investigation, it was found that the downcomer of the kerosene drawing area was plugged. All lines for the atmospheric distillation column were washed. Figure 19.147 shows the marked zones that indicate the time intervals during which the feed characteristic changes and oscillation occurs. When the top reflux amount rises above a certain value, the temperature fluctuation begins. During this oscillation, the change in pressure drop (ΔP) between the column top pressure and the flash zone pressure increases. This increase in ΔP showed the possible flooding inside the column.

The authors' findings of their first examination showed that with increased naphtha content in the processed crude, the amount of overhead reflux increases to keep the overhead temperature constant. This caused fluid head to form in the area where flooding was suspected. The liquid level formed exceeded the liquid carrying capacity of the tray at some point, causing it to fall into the lower trays, which makes them to cool. This incident creates the oscillation in the product withdrawal temperatures.

Pressure gauges were installed in different zones of the column (Figure 19.148) and pressure survey was conducted to determine the region where flooding occurs. The change in crude oil content may result in an increase in the amount of overhead reflux, depending on the ratio of the naphtha cut in the feed. Figure 19.149 shows the change in pressure drop with respect to the overhead reflux in the column.

Ozturk *et al.* [237] observed that when the overhead reflux exceeds a certain level, the difference between the column overhead pressure and the flash zone pressure reached ~ 370 g/cm^2, indicating that the liquid level accumulated in the column resulted in an excessive ΔP, since the theoretical and anticipated ΔP is around 200 – 300 g/cm^2. Figure 19.149 shows the change in ΔP on the column top and from the pressure gauge installed on the kerosene stripper with the amount of reflux. As the top reflux amount rises, the difference between the column top and the kerosene stripper pressure reaches 200–220 g/cm^2. Considering the number of trays in this region, these values correspond to three to four times the expected ΔP value, indicating that the suspected flooding in the column is in the region between the kerosene withdrawal zone and the top line of the column.

After the columns were steamed out and washed, commencing from the top of the atmospheric column, and further investigation was undertaken, a blockage was detected on the downcomer of the previous tray of the kerosene draw tray (Figure 19.150). A sample was collected to determine the source of the contamination. The kerosene upper tray is where the kerosene stream should flow from the relevant downcomer to the chimney tray's canal where the kerosene stream is drawn from the column. The contamination was the cause for the liquid level accumulation that resulted in the flooding. Figure 19.151 showed the fouling sample taken from the downcomer, and analysis further revealed 77% inorganic and 17% organic matter. The organic part was related to charge content and the inorganic part resulted from the accumulation in the kerosene downcomer. The inorganic matter showed that the pollution

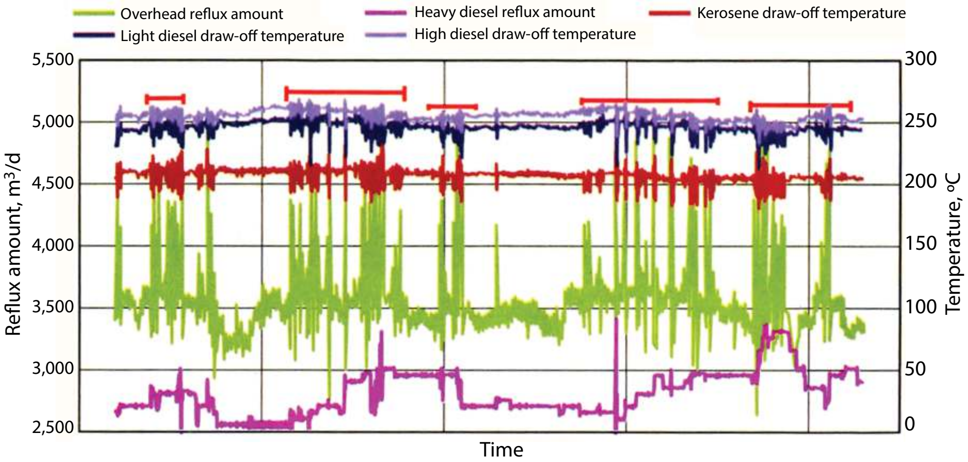


Figure 19.147 Reflux amounts and the oscillations in product draw-off temperatures (Source: Ozturk, G., *et al.* Hydrocarbon Processing, Feb. 2021 [237]).

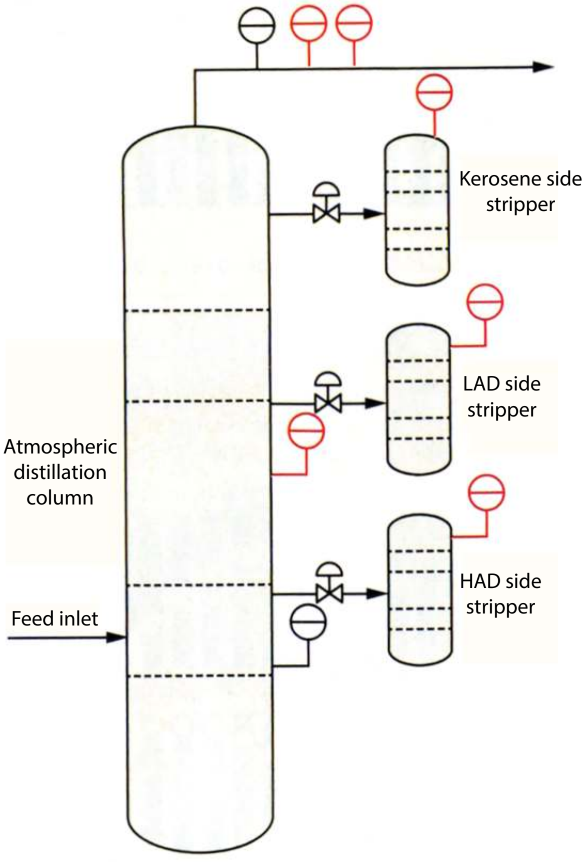


Figure 19.148 Locations of the installed pressure gauges (in red) and transmitters at the top and flash zones (Source: Ozturk, G., *et al.* Hydrocarbon Processing, Feb. 2021 [237]).

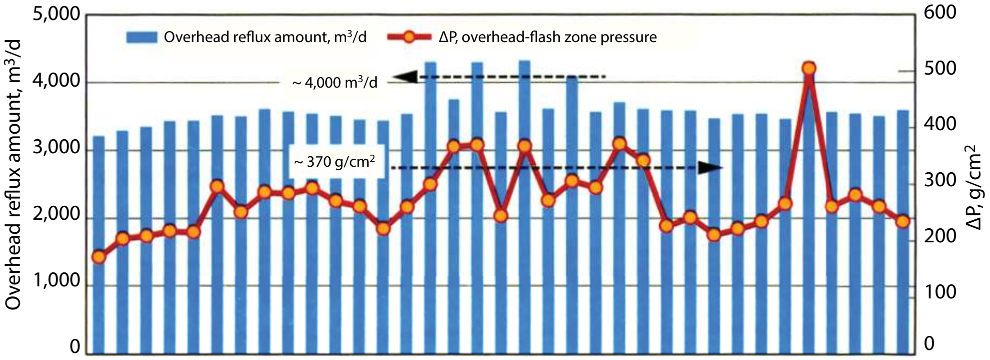


Figure 19.149 The ΔP through the column and the overhead reflux amount (Source: Ozturk, G., *et al.* Hydrocarbon Processing, Feb. 2021 [237]).

consisted of Fe, S and Ni elements. The root cause of the pollution in the downcomer was determined to be the accumulated corrosion pollution, which could not be inspected during the turnaround, since the upper tray would require to be removed.

Thorough cleaning was employed on the tray and canal after a water wash. The internals and trays of the column have been controlled since the cleaning, and no other major blockage or deposit has been detected. Figure 19.152 showed the flow path of the downcomer connection.

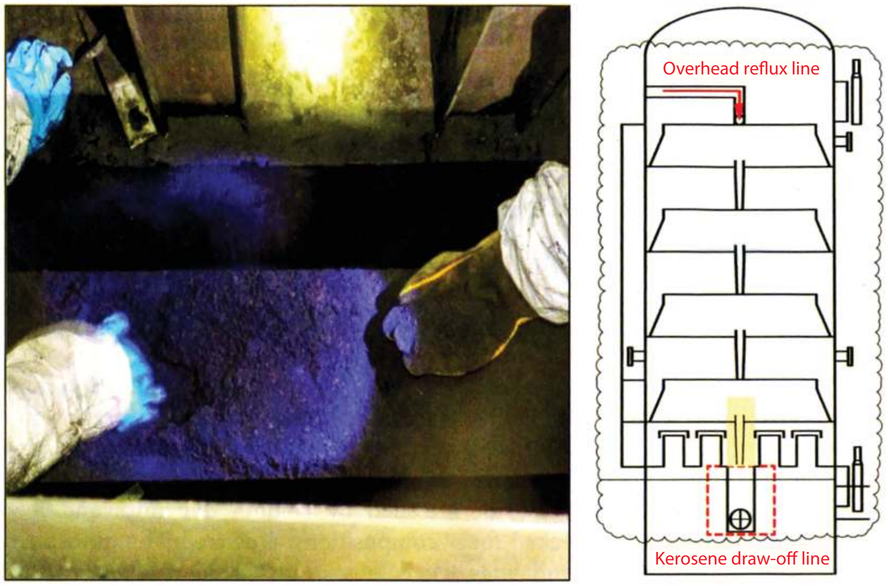


Figure 19.150 Blockage in the downcomer of the kerosene tray (Source: Ozturk, G., *et al.* Hydrocarbon Processing, Feb. 2021 [237]).

Figure 19.151 Accumulated fouling in the kerosene downcomer (Source: Ozturk, G., *et al.* Hydrocarbon Processing, Feb. 2021 [237]).

This case study illustrates the importance of cleaning the column and to maintain steam – out during emergency shutdown of the unit to guarantee operational efficiency after maintenance is carried out. Furthermore, the control of the column and its internal is essential; the cleaning of the downcomer sections and drawing lines must be provided, and the drawing lines must be washed until every line is cleaned.

Case Studies [204]

The following are problems that occur in some refinery facility units and how they were resolved.

Case Study 1

This problem occurred in a refinery crude unit after capacity augmentation. The flash point of heavy naphtha side cut 230–284 °F (110–140°C) was not coming up to the required levels despite adequate stripping inside the stripper of a CDU.

This is likely to occur in revamp applications where the downcomer area has not been increased and does not meet the usual for minimum residence time in the downcomer. It is always advisable to ensure minimum residence time even if the velocity in the downcomer and downcomer backup is adequate. In the shutdown of the unit, modification was carried out to increase the downcomer area, which resolved the problem.

Case Study 2

Proprietary Z bars are employed to increase the downcomer area without cutting and removing the existing bolting bars and rewelding the new bolting bars in revamp applications. This type of fitting reduces the work done inside the column to increase the downcomer area. A Z bar is fitted (without welding) on the existing bolting bars and a downcomer apron is fitted on it, increasing the downcomer area and weir length. The reduction in bubble area on the tray deck is compensated by increasing the number of dispersers on the active area. The capacity of a large naphtha splitter in a refinery was augmented by around 40% by fitting Z bars to increase downcomer area without cutting the existing bolting bars. The splitter was deployed for production of Isomerization (ISOM) feed and the continuous catalyst regeneration unit (CCRU) feed. Figure 19.153 shows a schematic drawing of Z bars.

Case Study 3

The crude distillation column of an atmospheric VDU was facing a problem since the capacity revamp of the unit: the required aircraft turbine fuel (ATF) pump-around (PA) flow was not achieved, and any attempt to increase the flow was resulting in loss of pump suction.

There was a packing bed for the pump-around, and after cooling, the pump-around liquid was returned at the top of the bed with a distributor. Due to the lower ATF pump-around, the cooling load was shifting to the top condenser (air cooler). Thus, the vapor velocity through the overhead air cooler was becoming alarmingly high and resulted in the thinning of the air cooler tubes. Furthermore, due to low pump-around flow and duty, less heat was transferred to the crude feed, resulting in low feed preheat. ATF yield was also low. Tray configuration was an overflowing and partial draw-off type (Figure 19.154).

Figure 19.155 shows the top view of the existing chimney orientation. The cause of the low ATF flow problem was narrowed down to faulty design in the ATF chimney tray, which was critically examined. There was inadequate cross-sectional area in the central collector channel, directing the liquids to draw-off sumps. There was also less area in the draw-off sump than necessary.

Figure 19.152 Tray downcomer connection and the flow path (Source: Ozturk, G., *et al.* Hydrocarbon Processing, Feb. 2021 [237]).

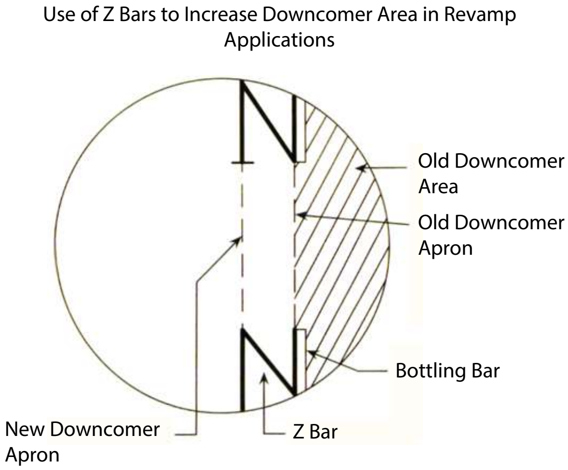


Figure 19.153 Tentative schematic drawing of Z bars [225].

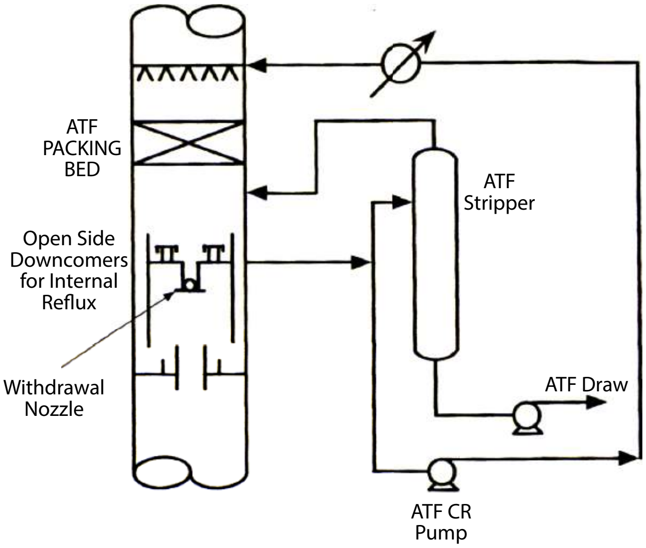


Figure 19.154 Existing withdrawal arrangement for ATF circulating reflux (CR) (Source: Ashis Nag, Distillation & Hydrocarbon Processing Practices, PennWell Co., 2016).

Withdrawal nozzle sizes were also found to be inadequate. The segmental weir below the packing did not have a hat. However, such a downcomer arrangement is theoretically correct if the liquid falling through the open segmental downcomers of the chimney tray is less than the desired internal reflux (IR).

It was checked and found that 10% of the liquid falling on the chimney tray from the packing above (which can drop through the open area of the downcomers) was less than the IR required for the tray below. Therefore, it should not require hats over the downcomers; however, there is likelihood of liquid channeling in the packing above cannot be eliminated. Due to channeling in the packing above, more than 10% of liquid can drop through the downcomers of chimney tray. This could result in a higher-than-desirable IR through the packing to the downcomers of the chimney tray and may result in additional reflux below the chimney tray. The additional IR in turn may result in more vapor condensation than desired in the tray below and result to lesser availability of ATF.

During the shutdown, the chimney tray was modified, as the chimneys are oriented transversely and the central collector is eliminated, diverting the liquids directly to the draw-off boxes. Also, pipe downcomers were installed in place of segmental downcomers. Hats are still provided on the top of the downcomers. The size of the withdrawal nozzles was increased to 18 in. to ensure self-venting flows (Figure 19.156).

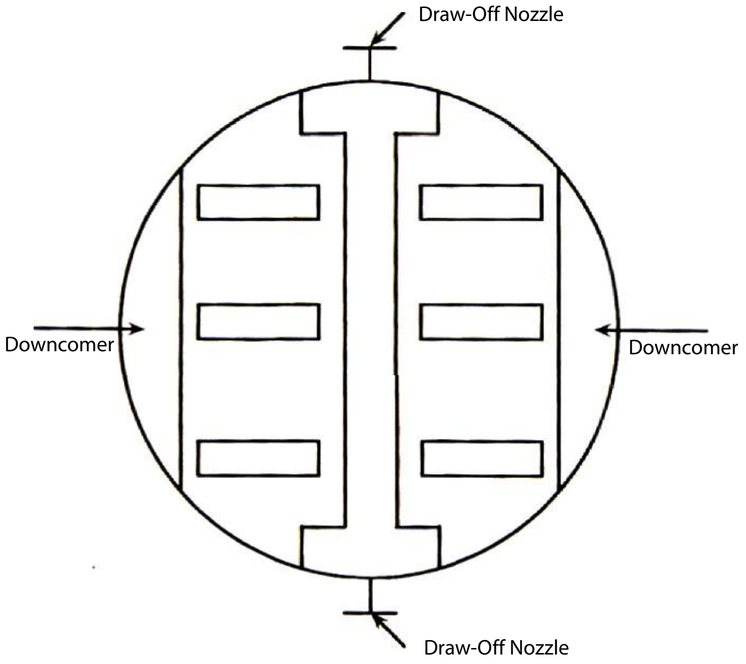


Figure 19.155 Top view of existing chimney orientation (Source: Ashis Nag, Distillation & Hydrocarbon Processing Practices, PennWell Co., 2016).

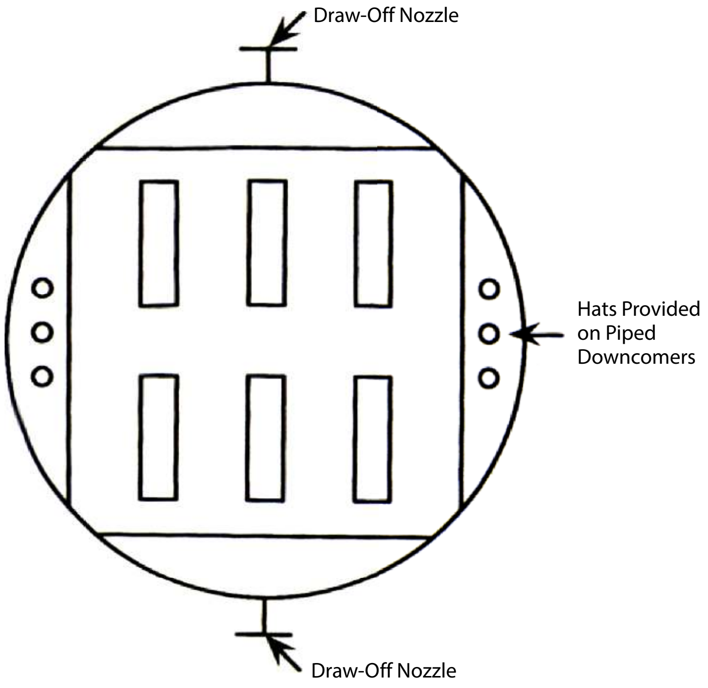


Figure 19.156 Modified chimney tray, draw-off box, and piped downcomers (Source: Ashis Nag, Distillation & Hydrocarbon Processing Practices, PennWell Co., 2016).

This problem was resolved as the pump-around flow increased to around 350 m^3/h, top reflux was reduced, preheat increased by 15°C, and the ATF yield increased by around 4% wt.

Case Study 4

The main fractionator of fluid catalytic cracking unit (FCCU) [which uses recycle oil (RCO) as feedstock] in a refinery unit was frequently experiencing high ΔP across the column during operation. The ΔP would increase quite quickly after startup of the unit.

It was reported that after the column was water-washed during shutdown, the ΔP would reduce after the startup and increase after a short period of operation. It was found that the performance of the desalter was not satisfactory. Thus, chlorides (Cl^-) in the feed can combine with ammonia (NH_3) and eventually deposit in the colder parts (top trays) of the column as ammonium chloride (NH_4Cl). To resolve the problem, an increase in column top temperature was first tried, but the refinery was unable to increase the fractionator top temperature due to the limitation of the endpoint of the FCCU gasoline. Therefore, the solution could not be continued. The refinery was advised to lower the pump-around duties to increase the top reflux ratio of the fractionator limited by the reflux drum temperature. This was suggested in order to increase the hydrocarbon partial pressure in the upper section of the column and thereby reduce the partial pressure of NH_3 and HCl, which would lower the deposition temperature of NH_4Cl. Also, the desalter operation was improved to arrest the chloride slippage, as the condition improved and the problem was resolved.

Case Study 5

A capacity revamp was carried out on a crude unit, which had a low pre-topper bottom residence time due to incorrect design. This was causing fluctuations in pre-topped crude flow through the downstream section during operation at higher capacity, which was in turn causing occasional low feed flow trips in the furnace. Low feed flow trips protect the furnace tubes from overheating and coking. The problem was resolved by providing an auxiliary vessel to improve the residence time (Figure 19.157).

Case Study 6

In a vacuum gas oil (VGO) hydrotreater, the stripper column preceding the fractionator was getting repeatedly choked. Visbroken naphtha along with the VGO was being routed for disposal. It was suspected that the deposition of primarily NH_4Cl (possibly part NH_4HS) on the upper (colder) part of the stripper was causing the choking of the dispersers of the fractionator. This was corrected by a modification in the upstream by adding an additional water injection at the inlet of the air cooler (located at the overhead of the hot low-pressure separator (HLPS)) for washing the NH_4Cl in the effluent stream. The unit already had a water-wash upstream of the cold high-pressure separator (CHPS). The problem was resolved by implementing the modification.

Case Study 7

In an FCCU main fractionator, frequent loss of bottom level was observed, with an increase in reactor temperature (increasing conversion). The bottom level was brought back to normal by decreasing the reactor temperature.

It was suspected that the main column bottom (MCB) pump-around return liquid (used to desuperheat the superheated feed vapor) might not be able to flow down through the openings of disc and doughnut trays, and was thus building up and not reaching the bottom, resulting in a lowering of the bottom level. The wash liquid below the heavy cycle oil (HCO) tray (wash zone) could also have been building up.

The vapor–liquid profile for the entire column was developed at highest conversion. This was given to the column internal designer to check the adequacy of disc and doughnut trays used at the desuperheating section. The designer ascertained that the disc and doughnut trays were not adequate to allow the flow of liquid and suggested changing them to grids and installed coarse packing in the HCO wash zone and sacrificing one wash-zone stage to allow the free flow of liquid. This resolved the problem as the capacity of the unit improved (Figure 19.158).

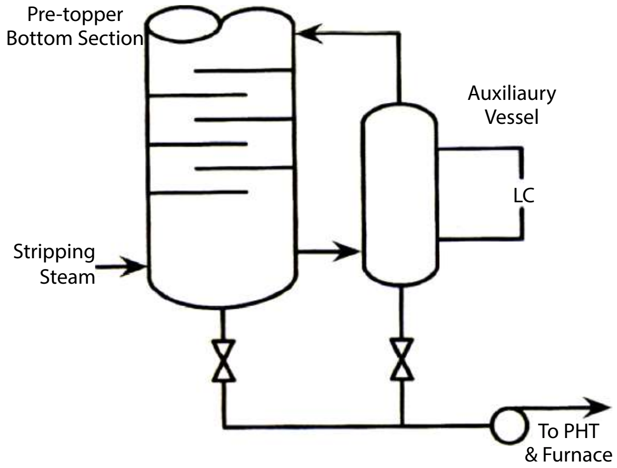


Figure 19.157 Use of auxiliary vessel to improve bottom residence time (Source: Ashis Nag, Distillation & Hydrocarbon Processing Practices, PennWell Co., 2016).

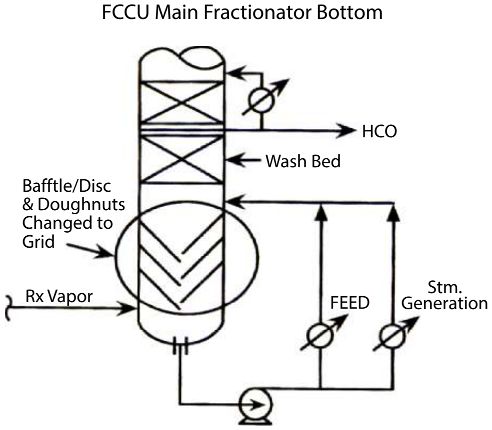


Figure 19.158 FCCU column desuperheating section.

19.54 Distillation Sequencing with Columns Having More than Two Products

In the separation of multicomponent mixture (more than two) using simple columns, there are only two possible sequences (see Figure 19.159). The sequence shown in Figure 19.159a is known as the direct sequence in which the lightest component is taken overhead in each column. The indirect sequence as shown in Figure 19.159b removes the heaviest component as bottom product in each column. If we consider the first characteristics of simple columns, a single feed is split into two products. One alternative to two simple columns is the possibilities shown in Figure 19.160. Here, three products are taken from one column. The designs are feasible and cost effective when compared with simple arrangements utilizing reboilers and condensers operating on utilities for certain ranges of conditions. Where the feed is dominated by the middle product (i.e., >50% of the feed) and the heaviest product is present in small quantities (i.e., <5%), then the arrangement shown in Figure 19.161a can be an attractive option. The heavy product must find its way down the column past the side stream. If the heavy product has a small flow and the middle product a high flow, a reasonably pure middle product cannot be achieved. In these situations, the side stream is usually taken as a vapor product to obtain a reasonably pure sidestream.

If the feed is dominated by the middle product (i.e., >50%) and the lightest product is present in small amounts (i.e., <5%), then the arrangement shown in Figure 19.161b can be an attractive choice. Here, the light product must find

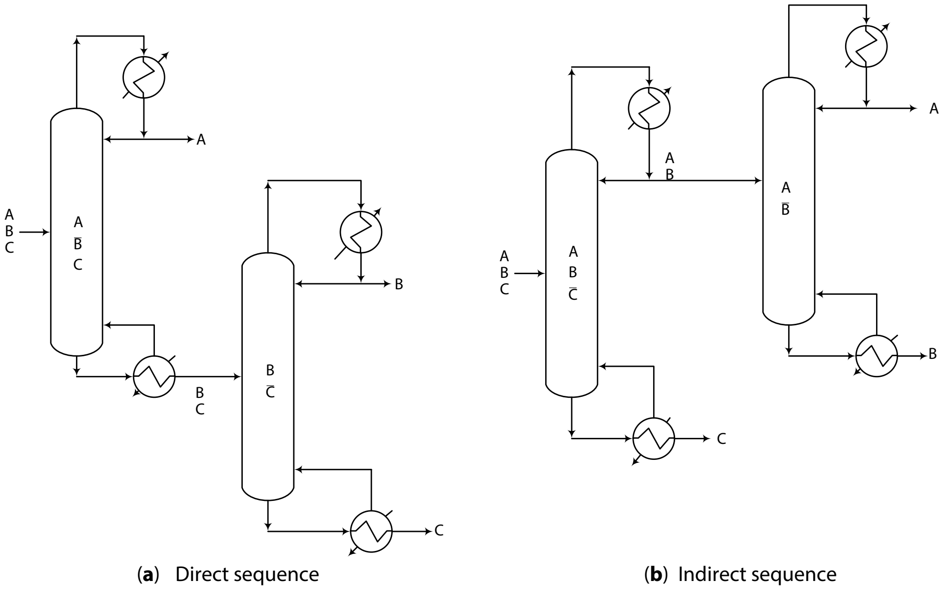
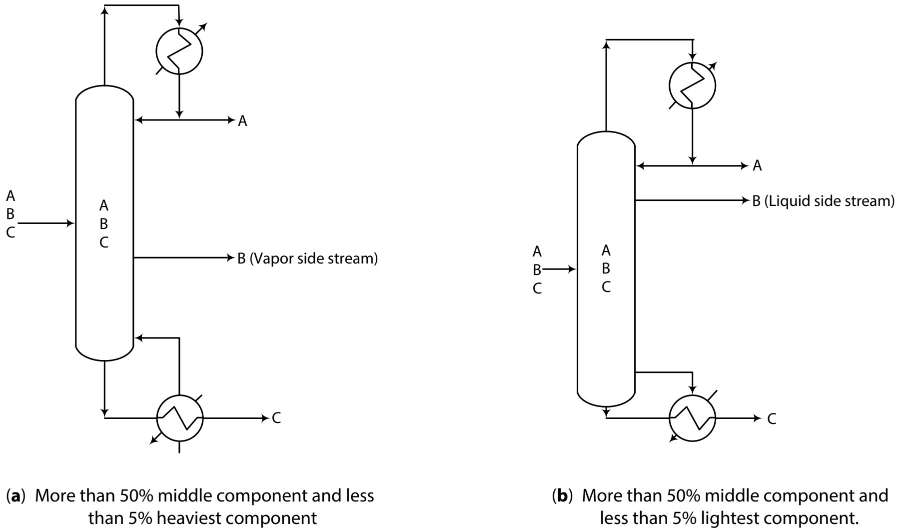


Figure 19.159 Alternative column sequences for a three-component separation.



(a) More than 50% middle component and less than 5% heaviest component

(b) More than 50% middle component and less than 5% lightest component.

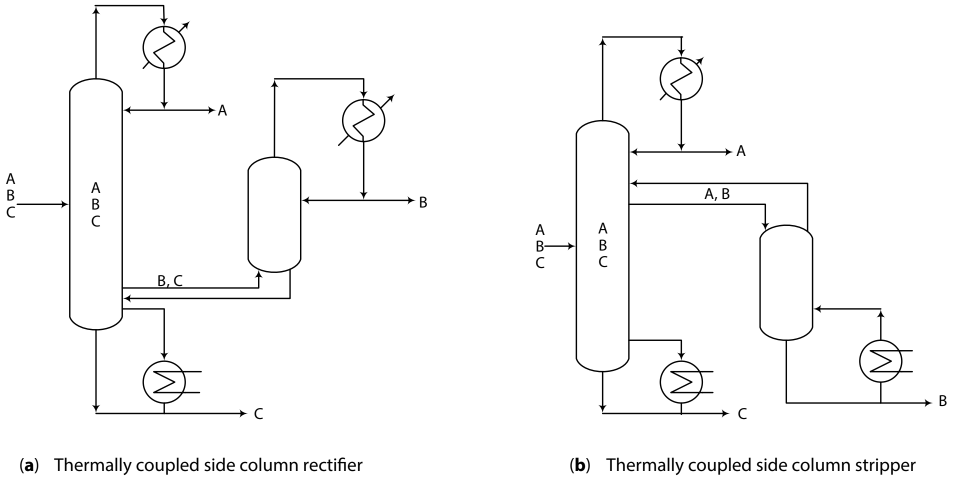
Figure 19.160 Distillation columns with three products (Source: Smith and Linnhoff, Trans. IChemE, ChERD, 66: 195, 1998).

its way up the column past the side streams. Unless the light product is a small flow and the middle product a high flow, a reasonably pure middle product cannot be achieved. In this instance, the side streams are taken as a liquid product to obtain a reasonably pure side stream [205].

Generally, single-column side stream arrangements can be attractive when the middle product is in excess and one of the other components is present only at small concentrations. Therefore, the side stream column only applies to special cases for the feed composition. More generally applicable arrangements are made possible by relaxing the restriction that separations must be between adjacent key components.

19.54.1 Thermally Coupled Distillation Sequence

The transfer of heat via direct contact is referred to as thermal coupling, and this is possible when materials flow in such a way as to provide some of the necessary heat transfer by direct contact. In the separation of multicomponent



(**a**) Thermally coupled side column rectifier (**b**) Thermally coupled side column stripper

Figure 19.161 Alternative column sequences for a three-component separation.

mixture, thermal coupled distillation columns require less energy and fixed cost as compared to conventional distillation of multicomponent mixture. To understand this process, first consider the separation of a ternary mixture having components **A**, **B**, and **C**. **A** is the most volatile component and **C** is the least volatile component, while component **B** has intermediate volatility. Conventionally, the mixture would be separated by the mixture either via direct sequence or indirect sequence as shown in Figures 19.159a and 19.159b.

For the given ternary mixture, choice between these two arrangements depends on boiling points, relative volatilities, composition of the ternary mixture, latent heat of vaporization, etc. By comparison, thermally coupled distillation columns need less energy and capital investment. Figures 19.161a and 19.161b show possible arrangements for the separation of the same ternary mixture of **A**, **B**, and **C** by thermally coupled distillation. In the thermally coupled side column rectifier, a vapor stream is withdrawn from the stripping section of the main column and sent for rectification in a side column with an overhead condenser. Liquid stream from the bottom of the side column is then returned to the main column.

In thermally coupled side column stripper, liquid stream from the main column is withdrawn and returned to the side column for stripping where the reboiler is provided. Vapor stream from side column is returned to the main column.

Selection between these two alternative arrangements depends on concentration profile of intermediate component **B**. If this reaches a maximum at any point of stripping section of the main or upstream column, then the side column rectifier is selected. The distillation of **B** from either **C** or **A** is made much easier by withdrawing a side stream from the point at which the concentration of **B** is maximized.

Figure 19.162 compares conventional and thermally coupled arrangements in terms of temperature and enthalpy. In the conventional arrangement, the pressure of the two columns can be set independently, therefore allowing variation in temperatures of the two condensers or the two reboilers. Such freedom does not exist in the thermally coupled arrangement. Although thermally coupled arrangement requires a smaller heat load than the conventional arrangement, more of the duties are at extreme levels. These smaller duties favor the benefit of heat integration, but the more extreme levels work to its detriment. Therefore, if a thermally coupled arrangement is to be integrated, then the smaller loads and the more extreme levels may work to advantage or disadvantage depending on the situation [206].

It is therefore recommended that simple columns should be used as a first option in the design process. Thermal coupling should be considered when the full heat integration context has been established.

An alternative to thermally coupled columns is a novel type of partitioned tray distillation column (also known as a divided-wall column or Petyluk column) for multicomponent system. Dividing-wall columns (DWCs) are a type

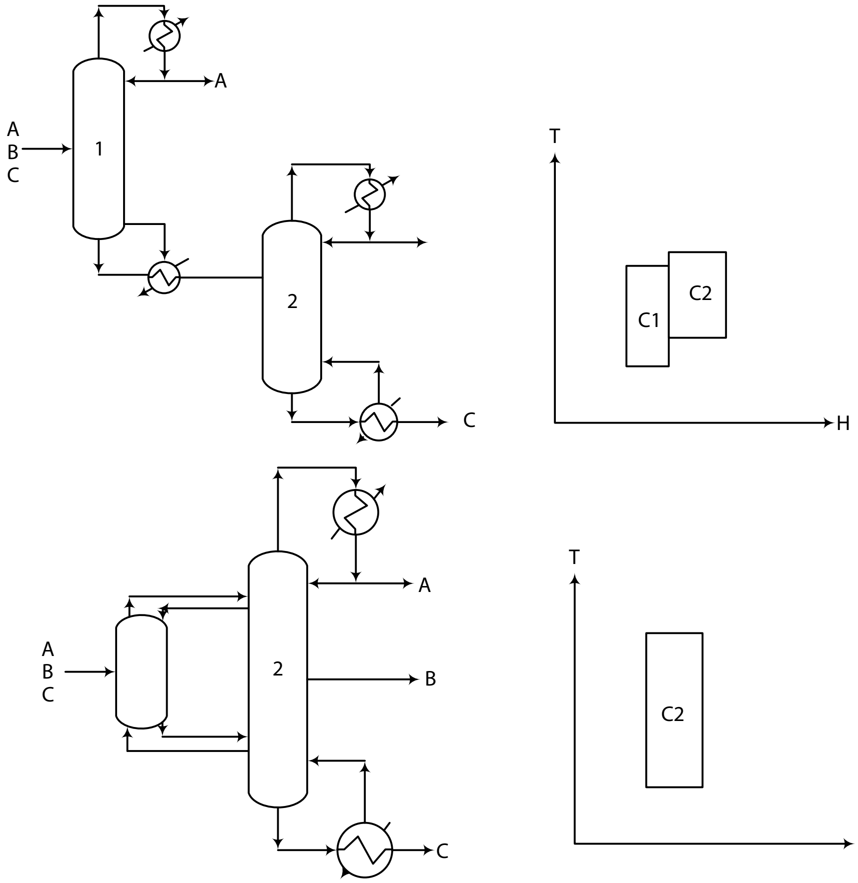


Figure 19.162 Relationship between heat load and level in simple and prefractionator sequences (Source: Smith and Linnhoff, Trans. IChemE, ChERD, 66: 195, 1998).

of distillation column that can separate mixtures of several components into three or more high-purity streams. A DWC requires much less energy, capital investment, and plant space than conventional columns in series or parallel configurations. DWCs have 20–30% lower capital cost requirements than conventional tower designs (Figures 19.163a and 19.163b).

Figure 19.163a shows such a column where a vertical baffle separates the feed location from that where intermediate boiling product is withdrawn. On the feed side of the partition, separation is achieved between the light (**A**) and heavy (**C**) fractions while the intermediate boiling fraction (**B**) is allowed to migrate to the top and bottom sides of the partition. On the other side of the partition, the light fraction flows down; so intermediate boiling fraction (**B**) at the desired purity is withdrawn. Such a column has a larger diameter, but overall capital and operating costs are significantly lower (~30 %) [205].

A Case Study [207]

A refinery wanted to upgrade its naphtha splitter to recover more medium-weight hydrocarbon. Figure 19.164 shows the original configuration of the naphtha splitter and depentanizer column at the facility.

The top product of the naphtha splitter consisted of about 15% nC_5 and 14% iC_5, with larger C_6 and C_7 components making up the remainder. This stream was routed to a depentanizer column, where the C_5 product and heavier naphtha components were separated.

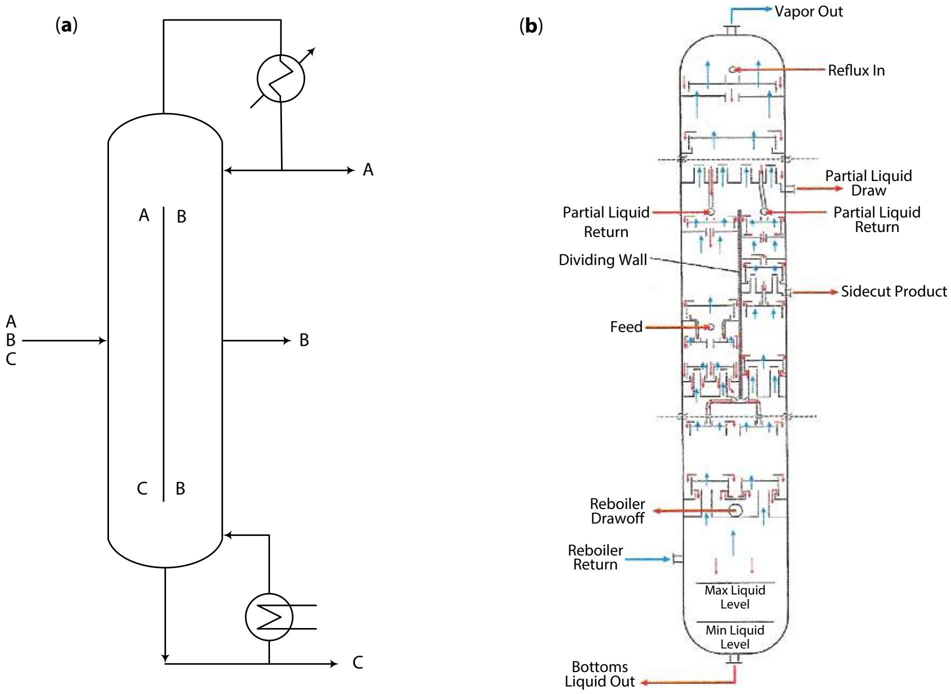


Figure 19.163 (a) partitioned distillation Column. (b) A standard dividing-wall column (DWC) has a wall that separates the column into two sections (Source: Manish Bhargava and Anju Patil Sharma, CEP, p. 30, Mar. 2019).

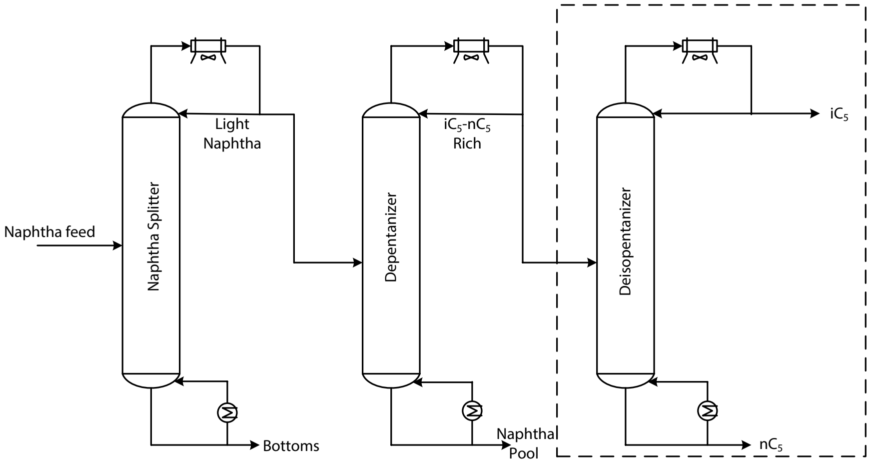


Figure 19.164 A new deisopentanizer installed downstream of the existing dependanizer column could be used to recover iC_5 as a separate product from nC_5 (Source: Manish Bhargava and Anju Patil Sharma, CEP, pp. 30–34, Mar. 2019).

To produce a premium gasoline product with a research octane number (RON) of 90, the refinery wanted to recover a stream of iC_5 with at least 90 wt% purity as a separate product. The refinery considered three possible solutions to achieve this product stream:

Option 1: Install a third column downstream of the existing depentanizer column.
Option 2: Revamp the depentanizer column by adding a side-cut stream.
Option 3: Revamp the depentanizer column to a DWC.

Option 1: Install a new deisopentanizer column

A new deisopentanizer column is built downstream of the original depentanizer column (Figure 19.164). The top product stream from the depentanizer, which is rich in iC_5 and nC_5, is sent to this deisopentanizer column, which further separates nC_5 to the column bottom and iC_5 to the top. Although this option meets the necessary product requirements, it requires substantial capital and operational expenses. A drawback of this option is that it has higher energy duties. The total required reboiler and condenser energy duties increase with an increasing number of columns in the sequence.

Option 2: Add a sidecut

This requires revamping the depentanizer column with a new column internals to add a side cut, which aids to produce iC_5-rich product at the top by separating a stream of mainly nC_5 from the middle section of the column (Figure 19.165). Since the side-cut column requires higher duty after the revamp, the column requires to be retrofitted with high-capacity trays. Tray spacing needs to be reduced to allow the column to accommodate more stages.

Remixing of components would be a problem in this conventional side-cut column, as this would reduce product quality. A better product purity could be achieved by increasing the reboiler duty. Furthermore, a compromise between product quality and duty could be attained by optimizing the location of the side cut to below the feed nozzle (Figure 19.166a). This would allow an iC_5-rich stream to be removed as the top product. However, a large amount of C_6 naphtha components would spill over to the side-cut streams, which would affect the quality of other two products. But if the location of the side cut was positioned above the feed nozzle (Figure 19.166b), then considerable iC_5 would mix and exit with the nC_5 stream.

Option 3: Add a dividing wall

In this option, the existing depentanizer is retrofitted with new column internals and converted to a middle DWC. The new column will produce high-purity iC_5 as the top product without requiring a significant increase in reboiler duty (Figure 19.167).

The middle DWC has three separation zones within a single shell. A common rectifying or enriching section is followed by a dividing wall section and a stripping section. The feed enters on the prefractionation side of the dividing wall. The lightest component in the feed mainly iC_5 travels up the column along with some nC_5. The heaviest components, i.e., the remaining nC_5, C_6, and C_7, travel down the column. In the common rectifying section, the majority

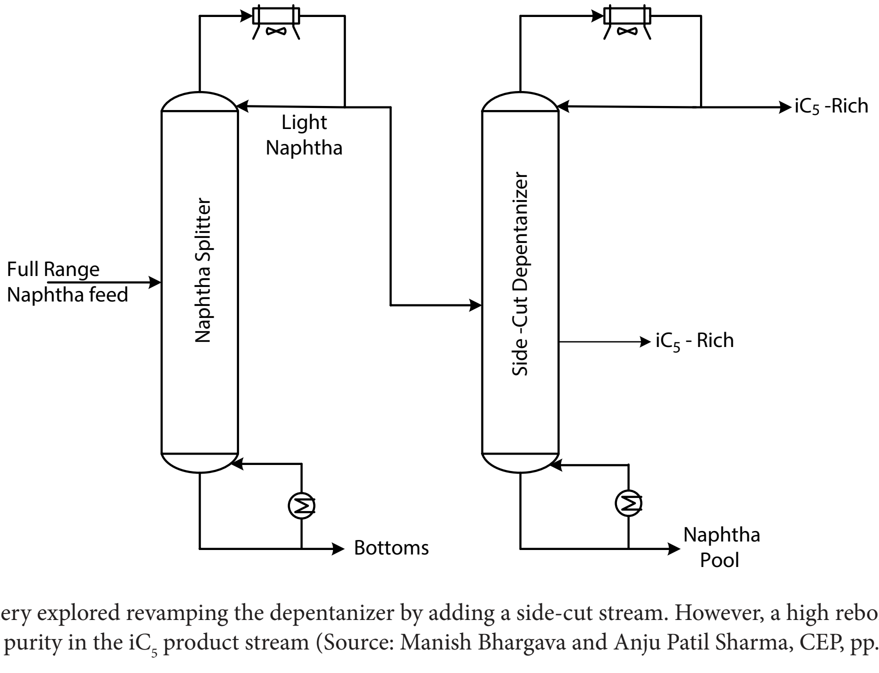


Figure 19.165 The refinery explored revamping the depentanizer by adding a side-cut stream. However, a high reboiler duty would be required to achieve high purity in the iC_5 product stream (Source: Manish Bhargava and Anju Patil Sharma, CEP, pp. 30–34, Mar. 2019).

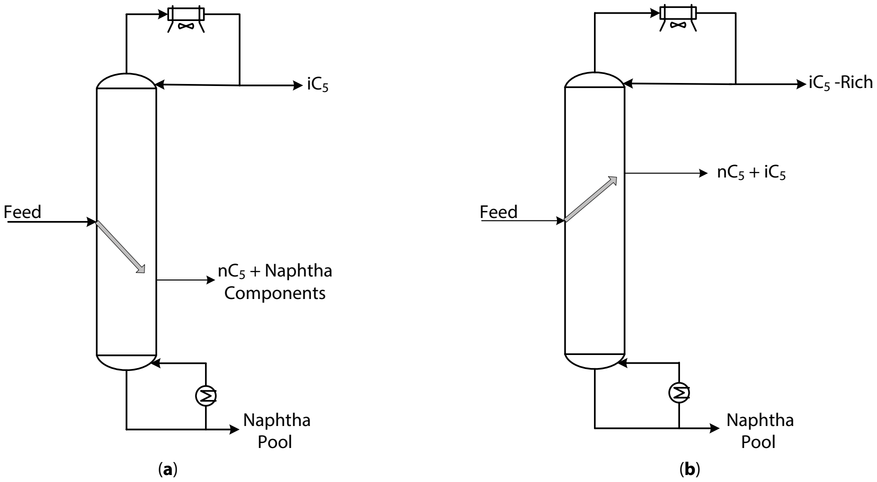


Figure 19.166 If the side cut of a conventional side-cut column is positioned below the level of the inlet feed, the top product will have a higher purity, but a considerable amount of naphtha heavy components will mix with the nC_5 side cut. Likewise, if the side cut is positioned above the level of the feed, a large amount of iC_5 will exit in the side-cut stream (Source: Manish Bhargava and Anju Patil Sharma, CEP, pp. 30–34, Mar. 2019).

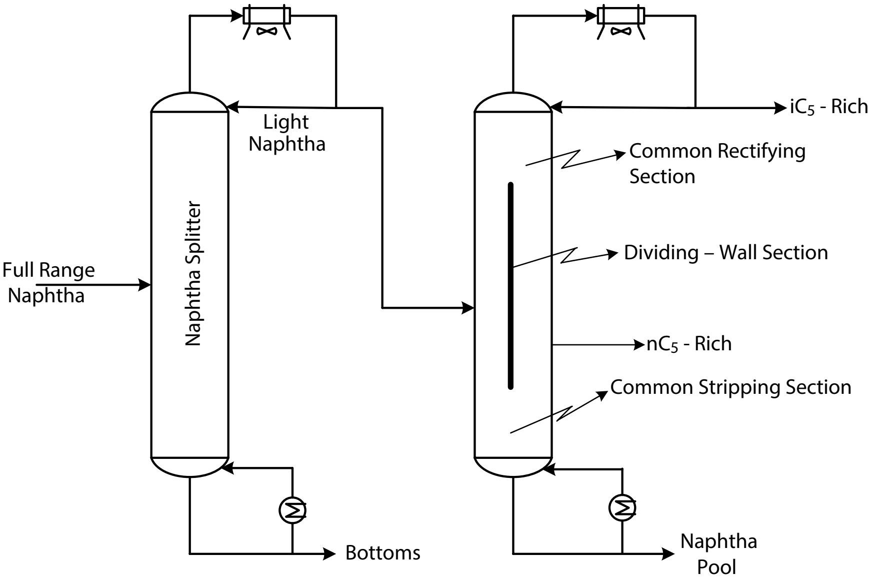


Figure 19.167 Adding a dividing wall to the middle of the depentanizer column creates an iC_5-rich product stream without requiring additional reboiler energy (Source: Manish Bhargava and Anju Patil Sharma, CEP, pp. 30–34, Mar. 2019).

of nC_5 flows down toward the dividing wall main fractionation side. High-purity iC_5 is recovered as the top product. Similarly, the nC_5 that flows with the heavier components travels up in the common stripping section. As a result, a concentration peak of nC_5 is achieved somewhere in the middle of the column on the product side of the dividing wall. A mixed C_5 stream is obtained as the side-cut product, while naphtha is removed at the bottom. Option 3 outperforms options 1 and 2 in equipment footprint and energy consumption as shown in Table 19.56.

Advantages and Disadvantages to Divided Wall Columns

Advantages

It has better energy efficiency and lower investment because multiple separations are done in one column. Also, it has lower capital and operating costs.

Table 19.56 A refinery compared three options to recover more medium-weight hydrocarbons. The DWC configuration outperforms the two-column sequence in equipment footprint and energy consumption [207].

Two-column sequence				
Criteria	**Dependanizer (Existing)**	**Deisopentanizer (New)**	**Side-cut column**	**Dividing wall column (DWC)**
Number of trays	50	75	75	75
Column diameter, m	4.6	3.7	4.6	4.6
Condenser duty, Million kcal/h	14.6	15.6	18.7	19.1
Total	30.2			
Reboiler duty, Million kcal/h	18.4	15.8	23.5	23.5
Total	34.2			

Disadvantages

The column is more complex and both design and control can be challenging. The entire possible range of feed and product compositions must be well understood. The equipment for controlling the liquid-vapor splits must be designed to accommodate the spectrum of composition and rate ranges. Alternatively, a design margin can be incorporated to deal with feed and product variations. However, allowing for extra reflux or extra boil up will reduce the capital and energy savings.

The total reboiler duty can decrease, but the duty at the highest temperature level may rise. This is because 100% of the reboiler duty is now at the equilibrium temperature and the heaviest product instead of being split across multiple products. In some heat integrated systems, this may offset much of the potential energy saving [238].

Further disadvantages are the control and design, which are more complex. As a result, troubleshooting may be more problematic and complicating because of more parameters that are involved [238].

19.54.2 Practical Constraints in Sequencing Options

The following constraints limit the design options that are practically feasible [208]:

1. Corrosive components must be removed early to minimize the use of expensive, corrosion resistant in materials of construction.
2. Reactive and heat-sensitive components must be removed early to prevent problems of product degradation.
3. Inhibitors are added to compounds that tend to polymerize when distilled. These tend to be nonvolatile, ending up in the column bottoms, and thus prevent finished products from being taken from the bottoms of columns.
4. Reactive and heat-sensitive components must be removed early to prevent problems of product degradation.
5. Safety considerations often dictate that a particularly hazardous component be removed as early as possible from the sequence in order to minimize the inventory of that material.
6. When components in the feed to a distillation sequence are difficult to condense, total condensation of these components might require low-temperature condensation employing refrigeration and/or high operating pressures, and hence increased operating costs. Therefore, the light components are normally removed from the top of the first column to minimize the use of refrigeration and high pressures in the overall sequence.

19.54.3 Choice of Sequence for Distillation Columns

Many heuristics/rules of thumb have been proposed for the selection of the sequence for distillation columns, and these are summarized in the following [208].

Heuristic 1. Separations where the relative volatility of the key components is close to unity or that exhibit azeotropic behavior should be performed in the absence of non-key components. That is, perform the most difficult separation last.
Heuristic 2. Sequences that remove the lightest components individually in column overheads should be favored, i.e., favor the direct sequence.
Heuristic 3. A component comprising a large fraction of the feed should be removed first.
Heuristic 4. Favor splits in which the molar flow between top and bottom products in individual columns is as nearly equal.

19.55 Heat Integration of Distillation Columns

Heat Integration in a Crude Distillation Unit

The design of a crude distillation unit (CDU) lies not only the desired products with the desired specifications but also one that will have efficient heat integration for achieving maximum heat recovery from hot products and result in high preheat and steam generation. Crude and vacuum distillation units (VDUs) are heat-surplus environments mainly because the heat supplied by the two furnaces is available to the crude.

Typically, the maximum preheat attainable with this configuration on a crude is around 563°F–572°F (295°C–300°C). The remainder of the heat from hot products is used for utility heating and steam generation. The low reduced crude oil (RCO) yield from the bottom of atmospheric CDU gives a lower preheat, but the recoverable heat from PA refluxes is often high, which provides an opportunity for the designer to configure the heat-exchanger train to recover the full potential of PA heat and attain high preheat. With overhead heat integration, it may be difficult to increase the preheat temperature by a substantial amount if the existing heat-exchanger train synthesis is very efficient. However, steam generation is possible using the hot streams which would have otherwise gone to heat the crude in the pre-desalter heat-exchanger train in the absence of overheat heat integration.

Column Overhead Heat Integration

Overhead integrating may or may not increase the crude preheat. It depends on the heat recovery pinch temperature. If the pinch temperature is higher, then the no-preheat gain will be achieved by integrating overhead heat, but the cooling duty for hot products will increase. However, additional steam generation may be possible with overhead integration, and this can be introduced only after a heat recovery pinch study and critical examination of the requirement have been carried out. Overhead heat integration can be used with the following options [225]:

1. Introduce a top PA in the topper.
2. Integrate pre-topper overhead heat with the crude if the pre-topper exists in the configuration.
3. Integrate main topper heat along with the crude.

Introducing a top PA is the easiest and cheapest way to integrate part of the overhead heat with the crude. This is achieved by reducing the reflux ratio, which lowers heat rejection to the overhead condensers or coolers. Some designs maintain the top temperature of the main fractionator solely that are a top PA and only have a provision for reflux as and when required as shown in Figure 19.168.

Pre-topper overhead integration is a good option and safe from the point of view of corrosion in the crude and overhead heat exchangers. Figure 19.169 shows a crude heat exchanger train of a pre-desalter heat exchanger where the pre-topper overhead is exchanging heat with the crude at a lower temperature. The first exchangers are pre-topper

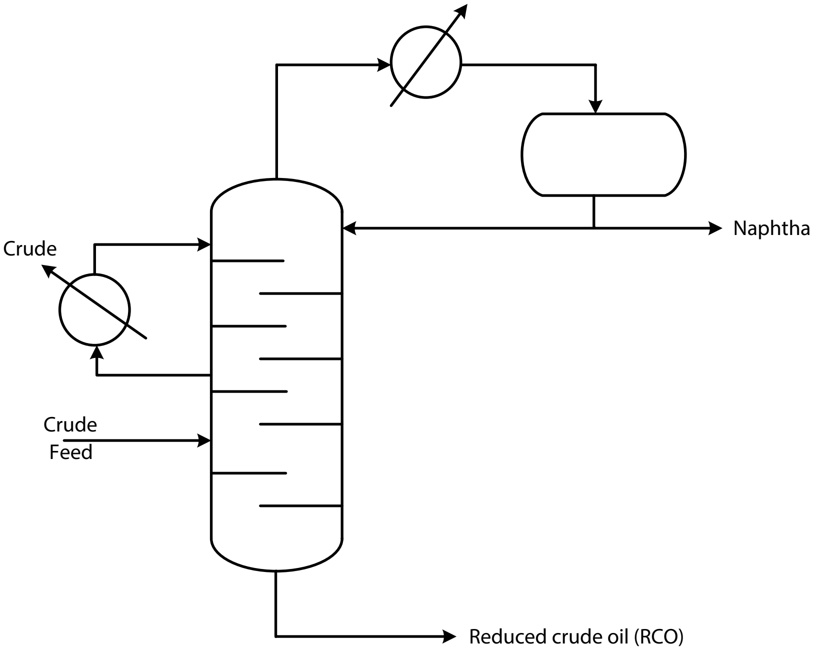


Figure 19.168 Basic scheme of overhead integration.

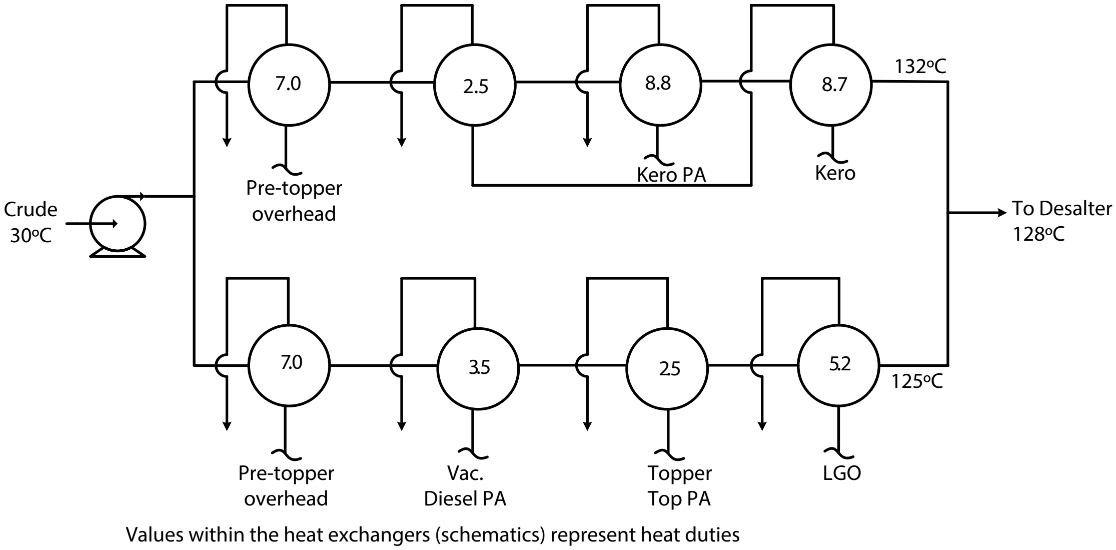


Figure 19.169 Pre-topper overhead coupled with crude for preheat.

overhead and crude. The pre-topper overhead is split and the crude is heated to 67°C from a temperature of 30°C with a heat transfer of around 14 Gcal/h of overhead heat to the crude. This also reduces the overhead trim condenser duty.

If the topper overhead is integrated, then it is likely to be coupled with the crude at almost the same location as the pre-topper overhead. If the overheads of both the topper and pre-topper are integrated with the crude, then there may be an excess of low-temperature heat.

If this is carried out, then some hot streams should be deployed to generate steam and then be routed to preheat train 1, so that the full benefit of overhead integration is utilized. Some steam generation using the light gas oil (LGO) stream can be carried out prior to routing this stream to the preheat train before the desalter. Other hot streams such as vacuum residue vacuum gas oil (VGO) and heavy vacuum gas oil (HVGO), which were likely to deliver heat in this train to maintain the desalter temperature can be used for the generation of steam and subsequently can be routed to the preheat train at lower temperatures. Therefore, the column overhead heat integration increases the efficiency of a unit through increased preheat, additional steam generation, or by preheating other cold streams. At locations where cooling water availability is a problem, lighter products that need cooling to 40°C–45°C are finally cooled with the cold crude or condensate (feed to the unit). The hot products first exchange heat with crude then are cooled with air coolers to around 65°C–700°C, followed by cold crude versus product exchangers in place of trim coolers (water coolers).

Integrated Atmospheric and Vacuum Distillation Units

Another method for heat integration is the use of atmospheric and vacuum units, where the reduced crude oil (RCO) is fed directly to the vacuum furnace before being fed to the vacuum distillation column. This is considered an energy efficient configuration, as no RCO heat is delivered to the air-cooler or the tempered water coolers and lost. Figure 19.170 shows a schematic of an integrated vacuum unit.

Generally, the scope for integrating conventional distillation columns into an overall process is limited due to practical constraints. The grand composite curve is the curve where enthalpy residuals are displayed as a function of the interval temperatures, and the enthalpy residuals corresponding to the highest and lowest interval temperatures are the minimum heating and cooling utility duties (Figure 19.151). This curve provides a quantitative tool to access the integrating viabilities of the column. If the column cannot be integrated with the rest of the process or if the potential for integrating is limited by the heat flows in the background process, then the distillation process and its complex arrangements should be reviewed.

The pinch in this context is the point of minimum temperature difference, representing a bottleneck in heat recovery (i.e., the point of closest approach of composite curves in a "heating and cooling") problem. It divides the process into two thermodynamically separate regions, above which only hot utility is required, and its magnitude corresponds to the overshoot of the cold composite curve (CCC). Below the pinch, only cold utility is necessary and its magnitude is given by the overshoot of the hot composite curve (HCC). Where the two composite curves overlap, the hot process streams are in enthalpy balance with the cold process streams. The appropriate placement of distillation columns is when heat integration is not across the pinch. In an inappropriately placed column, if it is shifted above the pinch

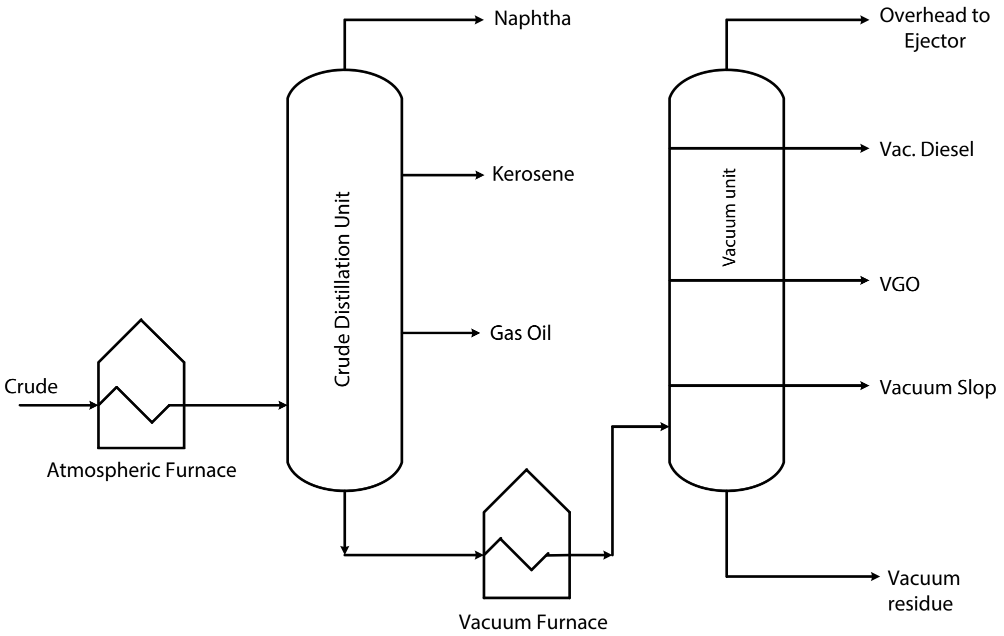


Figure 19.170 Integrated crude vacuum unit schematic.

by an increase in pressure, the condensing stream, which is a hot stream, is shifted from below to above the pinch. The reboiler stream, which is a cold stream, stays above the pinch. If the inappropriately placed column is shifted below the pinch by decreasing its pressure, then the reboiling stream, which is a cold stream, is shifted from above to below the pinch. The condensing stream stays below the pinch, and therefore appropriate placement is a case of shifting streams, which is a particular case of the plus/minus principle indicated by Smith and Linnhoff [206].

However, when a distillation column is inappropriately placed across the pinch, its pressure may be altered in order to achieve appropriate placement. However, as the pressure changes, the shape of the "box" changes, since not only do the reboiler and condenser temperatures change but also the difference between the two. This affects the relative volatility, generally decreasing with increasing pressure and subsequently both the height and width of the box will change as the pressure changes.

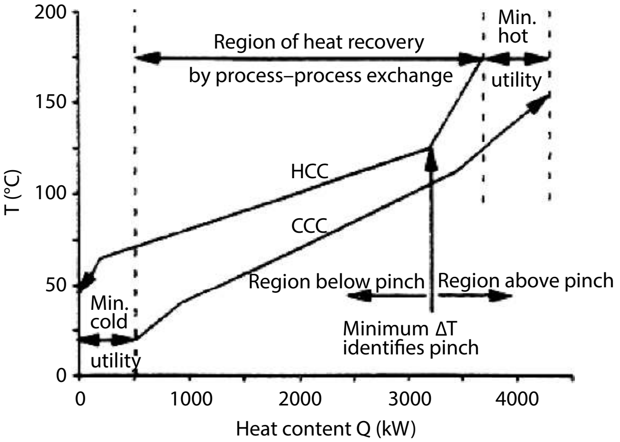
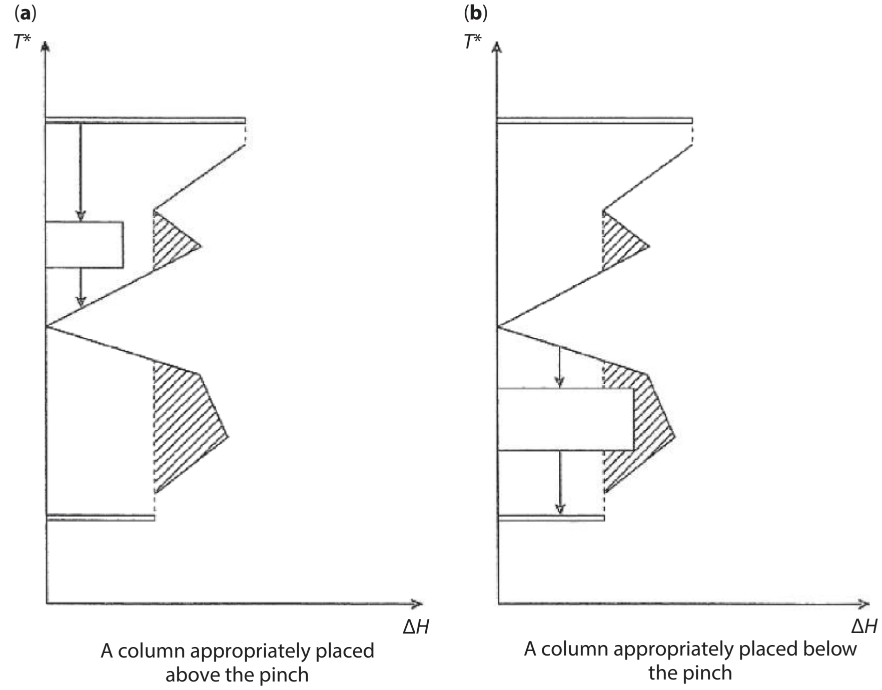


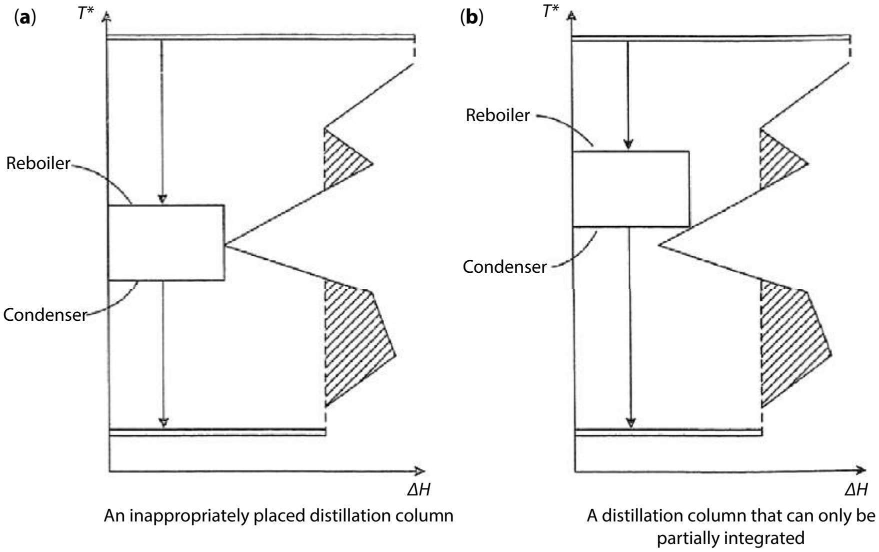
Figure 19.171 The hot composite cure (HCC) and cold composite curve (CCC) respectively show the heat availability and heat requirement for the overall process (Source: Shenoy, Udah, V., Heat Exchanger Network Synthesis: Process Optimisation by Energy and Resource Analysis, Gulf Publishing).



A column appropriately placed above the pinch

A column appropriately placed below the pinch

Figure 19.172 Distillation columns that fit against the grand composite curve (Source: Smith, R. and B. Linnhoff., Trans. IChemE, ChERD, 66, 195, 1988).



An inappropriately placed distillation column

A distillation column that can only be partially integrated

Figure 19.173 Distillation columns that do not fit against the grand composite curve (Source: Smith, R. and B. Linnhoff., Trans. IChemE, ChERD, 66, 195, 1988).

Changes in pressure also affect the heating and cooling duties for column feed and products. Therefore, the shape of the grand composite curve also changes as the column pressure changes. These effects will not be significant in most processes, since the sensible heat loads involved are small in comparison with the latent heat changes in condensers and reboilers. Figure 19.171 shows typical composite curves for both hot and cold streams indicating the minimum hot and cold utilities; the overlap between these curves indicates the maximum heat recovery, and ΔT is the pinch temperature. Figures 19.172 and 19.173 respectively show distillation columns that fit and do not fit against the grand composite curve. See Smith for detailed explanation [208].

If the distillation column will not fit either above or below the pinch, then other design options should be considered. One possibility is splitting the column feed and feeding to two separate parallel columns. The relative pressures in the columns are chosen such that the two columns can each be appropriately placed. The capital cost of such a scheme will be higher than that of a single column [206]. Use of an intermediate reboiler or condenser can also be considered. Reviews of this approach are presented elsewhere [209–211].

19.56 Capital Cost Considerations for Distillation Columns

Separators (e.g., distillation columns, evaporators, and dryers) are energy intensive. Their efficiency in terms of the overall process can be improved if they are properly heat integrated. In distillation, the major design parameters must be ascertained to allow the design to proceed. The first decision is the operating pressure. As this is raised:

1. Condenser temperature increases.
2. Vapor density increases resulting in a smaller column diameter.
3. Reboiler temperature increases with a limit often set by thermal decomposition of the material being vaporized. This results in excessive fouling.
4. Latent heat of vaporization decreases, i.e., reboiler and condenser duties become lower.
5. Separation becomes more difficult (relative volatility α decreases), i.e., more plates or reflux are required.

Correspondingly, as the operating pressure is lowered, these effects are reversed, and the lower limit is often set by the desire to avoid

- Refrigeration in the condenser.
- Vacuum operation.

The use of refrigeration and vacuum operation incurs further capital and operating costs and thus increases the complexity of the design. It is essential, therefore, to set distillation pressure to as low a pressure above ambient as allows cooling water or air cooling to be used in the condenser. The pressure should be fixed such that the bubble point of the overhead product is 10°C above the summer cooling water temperature or to atmospheric pressure if using vacuum operation. When distilling high molecular weight material, process constraints dictate that vacuum operation be used in order to reduce the boiling point of the material to below where product decomposition occurs.

Reflux ratio also needs to be chosen for distillation. Figure 19.174 illustrates a capital-energy tradeoff in a stand-alone distillation column. As the reflux ratio increases from its minimum, the capital cost decreases initially as the number of plates reduces from infinity, but the utility costs then increase as more reboiling and condensation are required. The optimal ratio of actual to minimum reflux is often less than 1.1, but designers are reluctant to design columns closer to minimum reflux than 1.1 except in special circumstances. This is because a small error in design data or small alteration in the operating conditions might result in an infeasible design. If the column is properly heat integrated with the rest of the process, the reflux ratio can often be increased without changing the overall energy consumption as illustrated in Figure 19.175a. Increasing the heat flow through the column decreases the number of plates required, but increases the vapor rate. The corresponding decrease in heat flow through the process (shown in Figure 19.175b) will have the effect of decreasing temperature driving forces and increasing the capital cost of the heat exchanger network. Therefore, the tradeoff for an appropriately integrated distillation column becomes one between the capital costs of the column and the capital cost of the heat exchanger network as shown in Figure 19.175c.

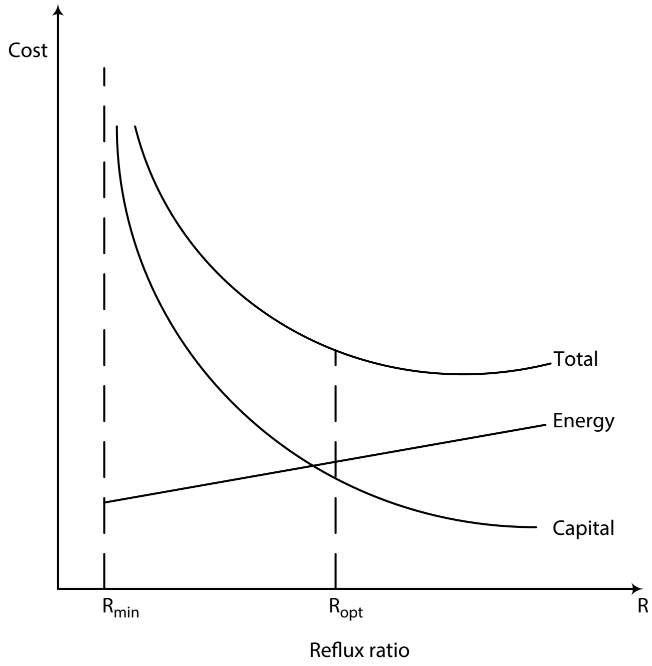


Figure 19.174 The capital energy tradeoff for stand-alone distillation column.

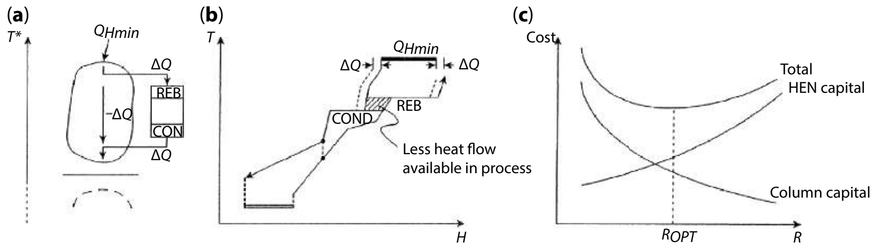


Figure 19.175 The capital/capital tradeoff for an appropriately integrated distillation column (Source: Smith and Linnhoff, Trans IChemE, ChERD, 66: 195, 1998).

The nature of the tradeoffs changes and the optimal reflux ratio for the heat-integrated column can be very different from that for a stand-alone column, and thus, the optimal reflux ratio for an appropriately integrated distillation column will be problem-specific.

Example 19.49

For the low-temperature distillation process shown in the flow diagram in Figure 19.176, calculate the minimum hot-utility requirement and the location of the heat recovery pinch. Assume the minimum acceptable temperature difference, ΔT_{min}, equals 5°C. The process streams passing through the seven heat exchangers appear in the first five columns of Table 19.57. The mass flow rates of the streams are being represented within the enthalpy (ΔH) values.

Solution

Calculation Procedure:

1. **For each stream, calculate its heat capacity flow rate**. For a given stream, the heat capacity flow rate CP is defined as ΔH divided by the absolute value of the difference between the supply temperature and the target temperature. For example, for Stream 1, the feed to Column 1, CP = 0.8/(20-0) = 0.04 MW/°C. The heat capacity flow rates for the seven streams appear in the last column of Table 19.58.
2. **For each stream, modify the supply and target temperatures to assure that the minimum temperature-difference requirement is met**. This step consists of lowering the supply and target temperature of each hot stream by $\Delta T_{min}/2$ and raising the supply and target temperatures of each cold stream by $\Delta T_{min}/2$. For stream 1 (a hot stream), for instance, the supply temperature drops from 20 to 17.5°C, and the target temperature drops 0 to -2.5°C; alternatively, for stream 4 (a cold stream), the supply temperature increases from 19 to 21.5°C, and the target temperature increases from 20 to 22.5°C. The shifted temperatures for the seven streams appear in Table 19.59.

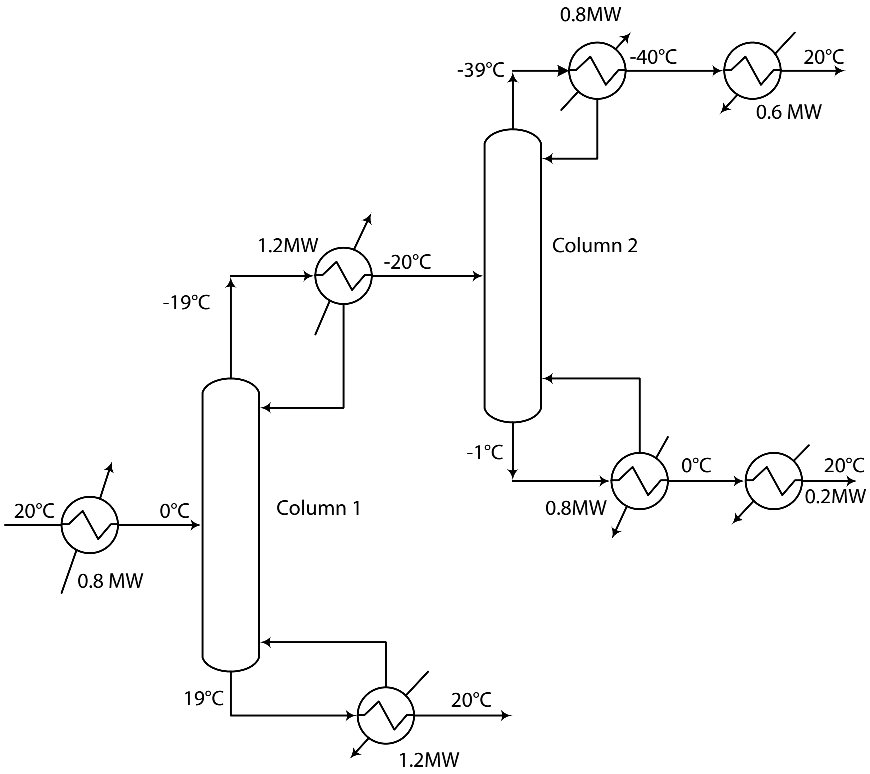


Figure 19.176 A low-temperature distillation process (Source: Smith, R., Chemical Process Design, McGraw-Hill, Co. 1995).

Table 19.57 Stream data for low-temperature distillation process.

Stream	Type	Supply temp. T_s (°C)	Target temp. TT (°C)	ΔH (MW)	Heat capacity flow rate, CP (MW °C^{-1})
1. Feed to column 1	Hot	-20	0	0.8	0.04
2. Column 1 condenser	Hot	-19	-20	1.2	1.2
3. Column 2 condenser	Hot	-39	-40	0.8	0.8
4. Column 1 reboiler	Cold	19	20	1.2	1.2
5. Column 2 reboiler	Cold	-1	0	0.8	0.8
6. Column 2 bottoms	Cold	0	20	0.2	0.01
7. Column 2 overheads	Cold	-40	20	0.6	0.01

Table 19.58 Streams names, supply temperatures and target temperatures with ΔT_{min} = 5°C.

Stream name	Supply temperature	Target temperature	dT min contrib	Heat capacity flowrate	Heat flow	Stream type	Supply shift	Target shift
	°C	°C	°C	kW/K	kW		°C	°C
Feed to column 1	20	0	2.5	40.000	800.0	HOT	17.5	-2.5
Column 1 condenser	-19	-20	2.5	1200.000	1200.0	HOT	-21.5	-22.5
Column2 condenser	-39	-40	2.5	800.000	800.0	HOT	-41.5	-42.5
Column 1 reboiler	19	20	2.5	1200.000	1200.0	COLD	21.5	22.5
Column 2 reboiler	-1	0	2.5	800.000	800.0	COLD	1.5	2.5
Column 2 bottoms	0	20	2.5	10.000	200.0	COLD	2.5	22.5
Column 2 overheads	-40	20	2.5	10.000	600.0	COLD	-37.5	22.5

3. Carry out a heat balance within each interval between the shifted temperatures. To display the temperature intervals, take all the 14 shifted supply and target temperatures (two for each of the seven streams) and list them in descending order (*noting that a few temperatures, such as 22.5°C, appear more than once, and thus do not repeat these*). The resulting list of stream names is the first column in Table 19.60. Draw horizontal lines extending leftward through each temperature value on the list. The regions between the lines are the temperature intervals, whose numerical values appear in the third column of Table 19.60. For example, the region between 21.5 and 17.5°C represents a temperature interval of (21.5–17.5) or 4°C.

Table 19.59 Shifted temperatures for the data in Table 19.57.

Stream	Type	Supply temp. T_s (°C)	Target temp. T_T (°C)	T_S^*	T_T^*
1	Hot	20	0	17.5	-2.5
2	Hot	-19	-20	-21.5	-22.5
3	Hot	-39	-40	-41.5	-42.5
4	Cold	19	20	21.5	22.5
5	Cold	-1	0	1.5	2.5
6	Cold	0	20	2.5	22.5
7	Cold	-40	20	-37.5	22.5

Then represent each of the seven streams by drawing vertical lines, starting at the shifted supply temperature and extending downward (for hot streams being cooled) or upward (for cold streams being heated) until reaching the shifted target temperature. For stream 1, for instance, the line begins at the line for 17.5°C and extends downwards to -2.5°C. Label each line with its CP value (e.g., CP = 0.04 for stream 1). The resulting seven vertical stream lines as shown in Figure 19.177 are referred to as streams and temperature intervals.

For each of the 10 temperature intervals, sum the heat capacity flow rates for the hot stream being cooled that falls within the interval, then subtract the sum of the heat capacity flow rate of each cold stream being heated that falls within the interval. The resulting algebraic sums are in the fourth column of Table 19.60. For example, there are three streams within the interval between 17.5 and -2.5 °C, namely Stream 1 being cooled with CP = 0.04, Stream 6 being heated with CP = 0.01, and Stream 7 being heated with CP = 0.01. The resulting algebraic sum is (0.04 – 0.01 – 0.01), or 0.02 as shown in Figure 19.157.

Finally, multiply each temperature difference in Column 3 by the corresponding algebraic sum of heat capacity flow rates in Column 4. The result is the heat balance or enthalpy change ΔH, in kilowatts for that interval. The intervals having a positive ΔH, such as the interval between 17.5 and -2.5°C, are designated as surplus, and the intervals with a negative ΔH are designated as deficit/demand. The results are shown in the fifth and sixth columns of Table 19.61.

4. **Cascade the surplus and deficit heat down the temperature scale from interval to interval, and note the largest net deficit that results. That number is the minimum hot utility requirement (Maximum energy recovery)**.

Cascaded heat is the amount of heat the problem has available from the hot streams over that required by the cold streams as we move from the higher temperatures to the lower ones. Anywhere we see a negative value in this cascaded heat column, we know that the hot streams have not produced enough heat to satisfy the needs of the cold streams above this entry. We look for the largest negative value or zero in this column and add this amount of heat Q_{min} as hot utility to the first interval in the cascade. All the net heat flows in the cascade now increase by this amount, and the minimum value becomes zero. This is the feasible heat cascade, or Problem Table (Table 19.61). The heat added to the first interval is the hot utility requirement (target) Q_{Hmin}, and the heat removed from the final interval is the cold utility target Q_{Cmin}. The point(s) at which there is zero net heat flow in the cascade is the pinch. The plot of the net heat flow (horizontal axis) against the shifted temperature (vertical axis) is the Grand composite curve (GCC).

Arrange the ΔH values as shown in Table 19.60, with the temperatures taken from the first column of the table. Assign a positive number if the cumulative value is surplus or a negative value if it is deficit. Starting from a value of 0.0, i.e., assuming no heat is supplied to the hottest interval 1 from hot utility,

Table 19.60 Temperature-interval heat balances for Example 19.49.

Shift temperature, °C	Interval	$T_{(i+1)} - T_i$ °C	mCp_{net} kW/K	$\Delta H_{INTERVAL}$ kW	Surplus/Deficit
22.5					
	1	1	-1220.0	-1220.0	Deficit
21.5					
	2	4	-20.0	-80.0	Deficit
17.5					
	3	15	20	300.0	Surplus
2.5					
	4	1	-770.0	-770.0	Deficit
1.5					
	5	4	30.0	120.0	Surplus
-2.5					
	6	19	-10.0	-190.0	Deficit
-21.5					
	7	1	1190.0	1190.0	Surplus
-22.5					
	8	15	-10.0	-150.0	Deficit
-37.5					
	9	4	0	0.0	----------
-41.5					
	10	1	800.0	800.0	Surplus
-42.5					

then the deficit of 1220 kW from interval 1 is cascaded into interval 2. There it joins the 80-kW deficit from interval 2, making a deficit of 1300 kW to cascade into interval 3. Interval 3 has a surplus of 300 kW; hence, after accepting the 1300 kW, it can be regarded as passing on a 1000-kW deficit to interval 4. Interval 4 has a 770-kW deficit and so passes on a 1770-kW deficit to interval 5 and so on, until it reaches interval 9 with 0 kW. A deficit of 800 kW leaves interval 9 and is cascaded into interval 10. Finally, the 800-kW surplus in interval 10 means that 0 kW is the final cascaded energy to the cold utility. Looking at the heat flows in Table 19.61, some are negative, which is thermodynamically infeasible as heat cannot be transferred up the temperature scale. To make the cascade just feasible, sufficient heat must be added from a hot utility to make the heat flow at least zero. The smallest amount of heat required from a hot utility is the largest negative heat flow from Table 10.61, i.e., 1840 kW. In Table 19.60, 1840 kW is added to the first interval from a hot utility. This does not change the heat balance within each interval but increases all the heat flows between the intervals by 1840 kW, giving one heat flow of just zero at an interval temperature of -21.5°C. To confirm this, repeat the downward cascading but begin with an assumed input of 1840 kW at the top of the cascade, as shown in Table 19.62. Note that, as a result, no deficit results. The top value in the last column of Table 19.62, 1840 kW, is the minimum amount of heat

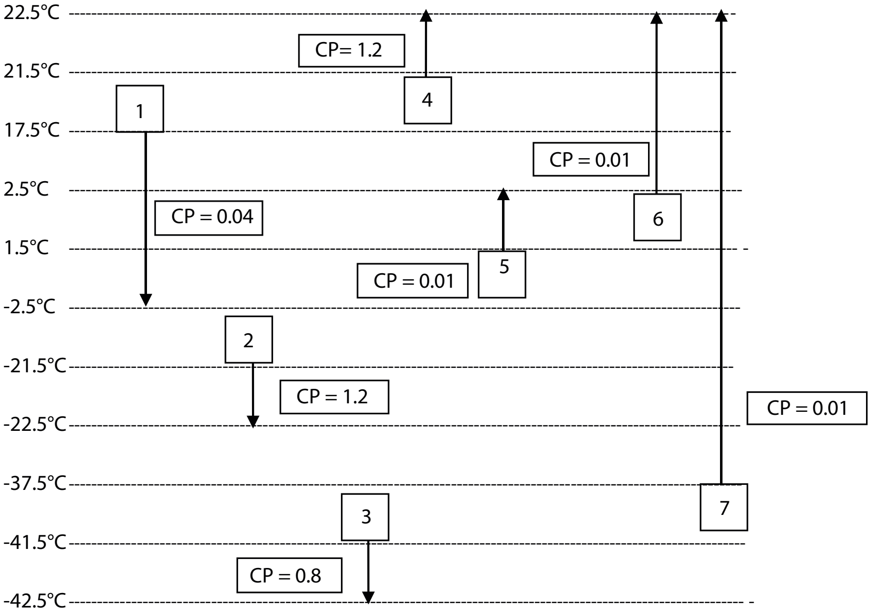


Figure 19.177 Streams and temperature intervals.

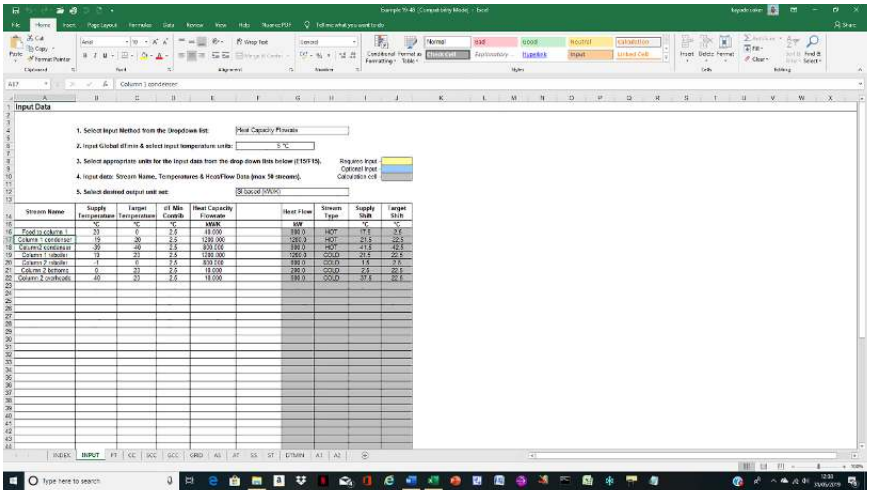
Table 19.61 Problem table and cascade.

Shift Temperature, °C	Interval	T(i+1) – Ti °C	(ΣCP – ΣCP) kW/°C	ΔH kw		Infeasible Cascade From Hot Utility		Feasible Cascade From Hot Utility	
22.5						⇩	0	⇩	1840
	1	1	−1.22	−1220		−1220		−1220	
21.5						⇩	−1220	⇩	620
	2	4	−0.02	−80		−80		−80	
17.5						⇩	−1300	⇩	540
	3	15	0.02	300		300		300	
2.5						⇩	−1000	⇩	840
	4	1	−0.77	−770		−770		−770	
1.5						⇩	−1770	⇩	70
	5	4	0.03	120		−120		−120	
−2.5						⇩	−1650	⇩	190
	6	19	−0.01	−190		−190		−190	
−21.5					PINCH	⇩	−1840	⇩	0
	7	1	1.19	1190		1190		1190	
−22.5						⇩	−650	⇩	1190
	8	15	−0.01	−150		−150		−150	
−37.5						⇩	−800	⇩	1040
	9	4	0	0		0		0	
−41.5						⇩	−800	⇩	1040
	10	1	0.8	800		800		800	
−42.5						⇩	0	⇩	1840
						To Cold Utility		To Cold Utility	

we must put into the problem from hot utilities; the bottom value, 1840 kW, is the minimum amount of heat we must remove from the problem using cold utilities. The Problem Table of Figure 19.159 shows that the hot stream pinch temperature = - 19°C and cold stream pinch temperature = - 24°C.

5. **Find the heat-recovery pinch point**. The pinch arises at the point where, after adding the minimum heat utility, the cumulative sum of ΔH values is zero. In this case, it occurs at -21.5°C. Assuming that the heat exchangers become networked appropriately, and keeping in mind that the pinch result of -21.5°C emerges after the temperatures were shifted, in step 2, we note that at the pinch, the hot stream being cooled is a temperature of $\left(-21.5+\Delta T_{min}/2\right)$ or (-21.5 + 2.5), or -19°C, and the cold stream being heated is at $\left(-21.5-\Delta T_{min}/2\right)$ or -24°C.

Figures 19.178 and 19.179 show the screenshots of the input data and the problem table, respectively. Figure 19.180 shows the screenshot of the plot of hot and cold composite curves, i.e., the actual temperature against the heat flow



Column 1 condenser

Input Data

1. Select Input Method from the Dropdown list: Heat Capacity Flowrate

2. Input Global dTmin & select input temperature units: 5 °C

3. Select appropriate units for the input data from the drop down lists below (E15/F15).

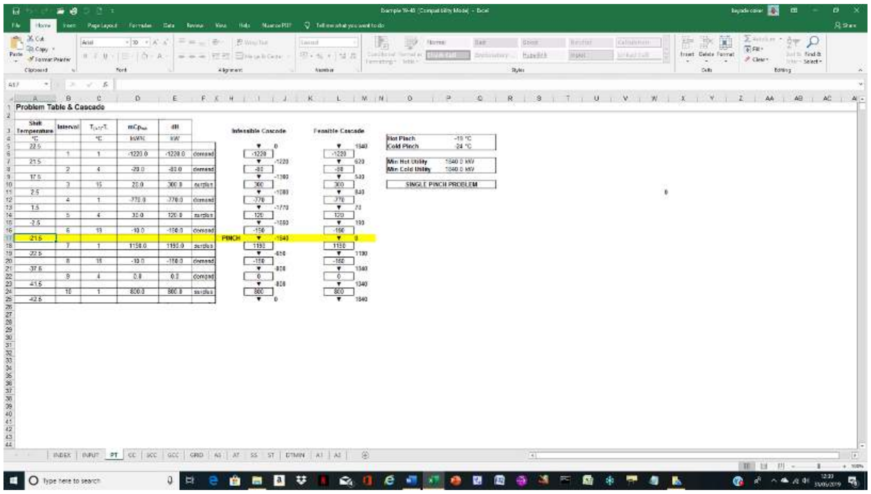
4. Input data: Stream Name, Temperatures & Heat/Flow Data (max 50 streams).

5. Select desired output unit set: SI based (kW/K)

Required input 🟨
Optional input 🟦
Calculation cell ▨

Stream Name	Supply Temperature	Target Temperature	dT Min Contrib	Heat Capacity Flowrate		Heat Flow	Stream Type	Supply Shift	Target Shift
	°C	°C	°C	kW/K		kW		°C	°C
Feed to column 1	20	0	2.5	40.000		800.0	HOT	17.5	2.5
Column 1 condenser	19	20	2.5	1200.000		1200.0	HOT	21.5	22.5
Column2 condenser	-39	-40	2.5	800.000		800.0	HOT	-41.5	-42.5
Column 1 reboiler	19	20	2.5	1200.000		1200.0	COLD	21.5	22.5
Column 2 reboiler	-1	0	2.5	800.000		800.0	COLD	1.5	2.5
Column 2 bottoms	0	20	2.5	10.000		200.0	COLD	2.5	22.5
Column 2 overheads	-40	20	2.5	10.000		600.0	COLD	-37.5	22.5

Figure 19.178 Snapshot of the input data of Example 19.49.



Problem Table & Cascade

Shift Temperature	Interval	$T_{(i+1)}$-T_i	mCp_{net}	dH			Infeasible Cascade		Feasible Cascade	
°C		°C	kW/K	kW						
22.5							▼	0	▼	1840
	1	1	-1220.0	-1220.0	demand		-1220		-1220	
21.5							▼	-1220	▼	620
	2	4	-20.0	-80.0	demand		-80		-80	
17.5							▼	-1300	▼	540
	3	15	20.0	300.0	surplus		300		300	
2.5							▼	-1000	▼	840
	4	1	-770.0	-770.0	demand		-770		-770	
1.5							▼	-1770	▼	70
	5	4	30.0	120.0	surplus		120		120	
-2.5							▼	-1650	▼	190
	6	19	-10.0	-190.0	demand		-190		-190	
-21.5						PINCH	▼	-1840	▼	0
	7	1	1190.0	1190.0	surplus		1190		1190	
-22.5							▼	-650	▼	1190
	8	15	-10.0	-150.0	demand		-150		-150	
-37.5							▼	-800	▼	1040
	9	4	0.0	0.0	demand		0		0	
-41.5							▼	-800	▼	1040
	10	1	800.0	800.0	surplus		800		800	
-42.5							▼	0	▼	1840

Hot Pinch -19 °C
Cold Pinch -24 °C

Min Hot Utility 1840.0 kW
Min Cold Utility 1840.0 kW

SINGLE PINCH PROBLEM

Figure 19.179 Snapshot of the problem table of Example 19.49.

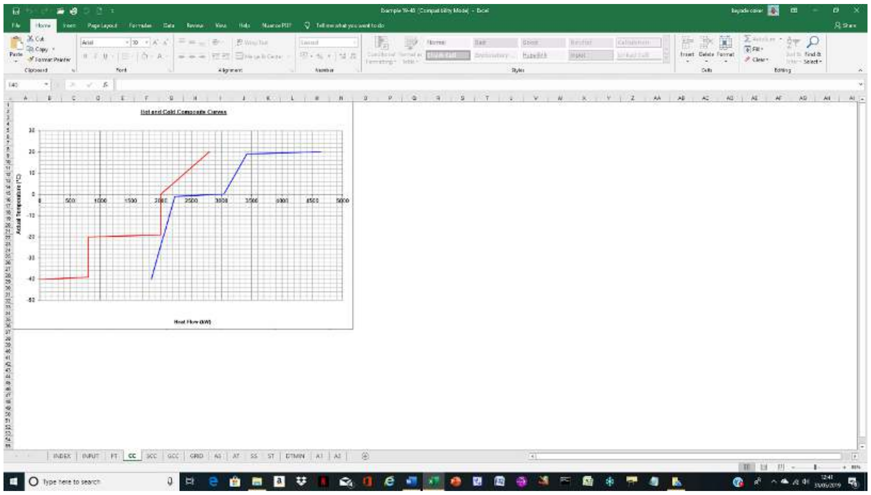


Figure 19.180 Snapshot of the hot and cold composite curves of Example 19.49.

(ΔH) where individual cold streams are combined into a composite cold stream and individual hot streams are combined into a composite hot stream, and Figure 19.181 shows the screenshot of the shifted hot and cold composite curves. Figure 19.182 shows the screenshot of GCC of the shifted temperature against the net heat flow, respectively. From Figure 19.183, the minimum heat and cold utility requirements (or maximum energy recovery) at $\Delta T_{min} = 5°C$ can be determined. Figure 19.179 shows the screenshot of the problem table with the pinch temperature of −21.5°C and Figure 19.183 represents the difference between the heat available from the hot streams and the required heat by the cold streams, relative to the pinch, at a given shifted temperature. Thus, the GCC is a plot of the net heat flow against the shifted (interval) temperature, which is a graphical plot of the problem table (heat cascade) of Tables 19.57 and 19.58, respectively. The GCC not only tells us how much net heating and cooling is required; it also tells us what temperatures it is needed at. There is no need to supply all the utility heating at the highest temperature interval; much of it can, if desired, be supplied at lower temperatures. As an additional check, the cold utility target minus the hot utility target should equal the bottom value of the infeasible heat cascade, which is 0 kW as shown in Table 19.60.

Note: In earlier papers, Equation (19.516)

$$\Delta H_i = (S_i - S_{i+1})\left(\sum CP_H - \sum CP_C\right)_i \tag{19.516}$$

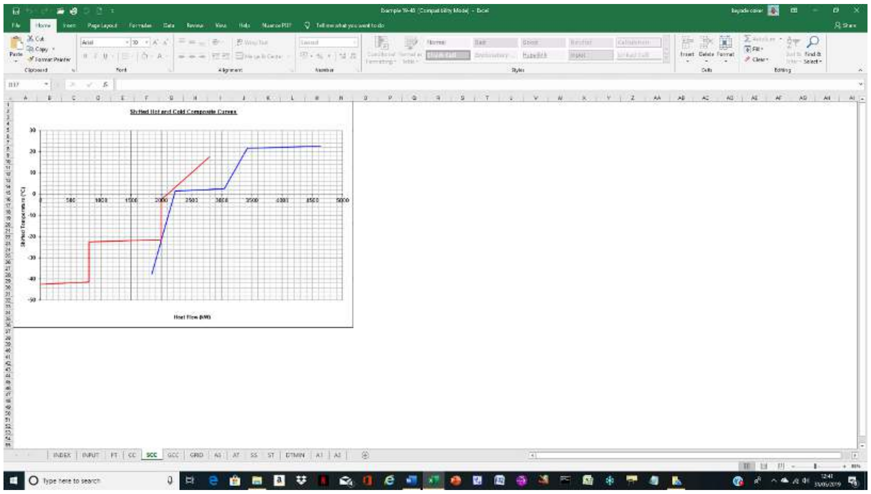


Figure 19.181 Snapshot of the shifted hot and cold composite curves of Example 19.49.

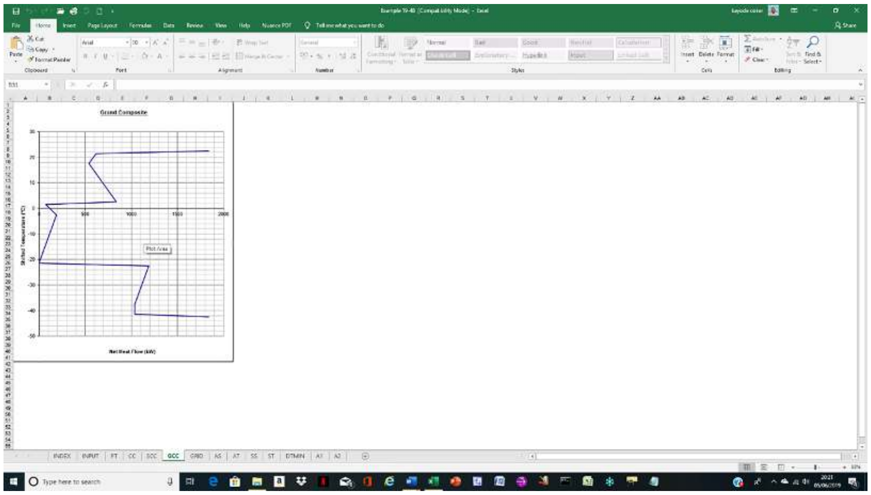
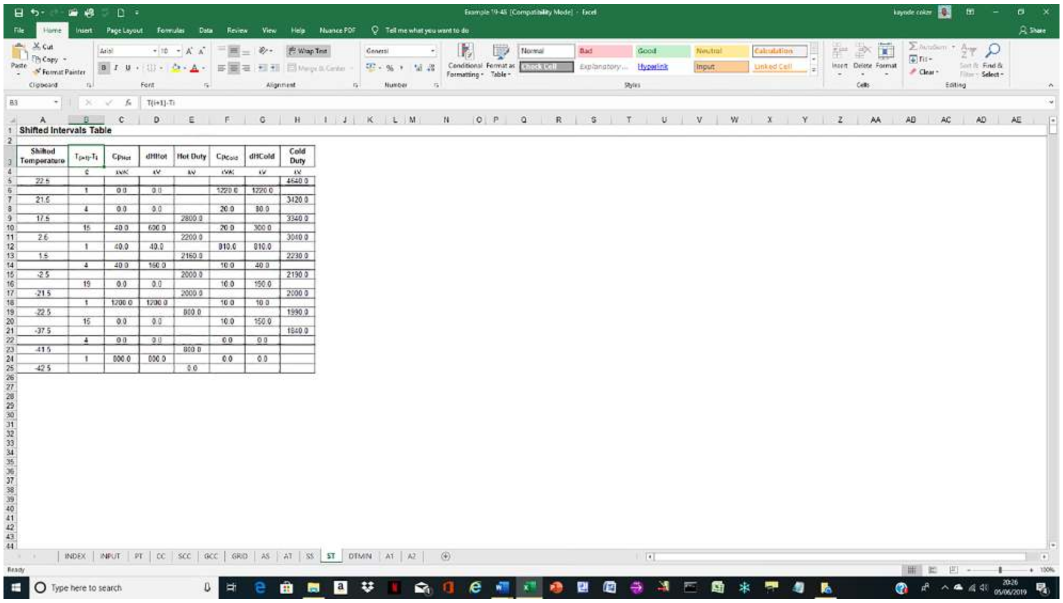


Figure 19.182 Snapshot of the grand composite curve of Example 19.49.



Shifted Intervals Table

Shifted Temperature	$T_{(i+1)}-T_i$	Cp_{Hot}	dHHot	Hot Duty	Cp_{Cold}	dHCold	Cold Duty
	C	kW/K	kW	kW	kW/K	kW	kW
22.5							4640.0
	1	0.0	0.0		1220.0	1220.0	
21.5							3420.0
	4	0.0	0.0		20.0	80.0	
17.5				2800.0			3340.0
	15	40.0	600.0		20.0	300.0	
2.5				2200.0			3040.0
	1	40.0	40.0		810.0	810.0	
1.5				2160.0			2230.0
	4	40.0	160.0		10.0	40.0	
-2.5				2000.0			2190.0
	19	0.0	0.0		10.0	190.0	
-21.5				2000.0			2000.0
	1	1200.0	1200.0		10.0	10.0	
-22.5				800.0			1990.0
	15	0.0	0.0		10.0	150.0	
-37.5							1840.0
	4	0.0	0.0		0.0	0.0	
-41.5				800.0			
	1	800.0	800.0		0.0	0.0	
-42.5				0.0			

Figure 19.183 Snapshot of shifted interval table of Example 19.49.

was reversed; temperature intervals with a net deficit were shown as positive and with a net surplus as negative. However, this is counter-intuitive [212] and more recent practice has been for a heat surplus to be positive.

where
ΔH = heat content
S = shifted temperature
CP_H = heat capacity flow rate of hot fluid
CP_C = heat capacity flow rate of cold fluid

Since no unit should have a temperature difference smaller than ΔT_{min}, a consequence is such that there should be no heat transfer across the pinch. There are three possible ways of moving the hot and cold composite curves closer together by ΔT_{min} so that they touch at the pinch, and this may be achieved in three ways.

1. Express all temperatures in terms of hot stream temperature and increase all cold stream temperatures by ΔT_{min}.
2. Express all temperatures in terms of cold stream temperatures and reduce all hot stream temperatures by ΔT_{min}.
3. Use the shifted temperatures, which are a mean value; all hot stream temperatures are reduced by $\Delta T_{min}/2$ and all cold stream temperatures are increased by $\Delta T_{min}/2$.

The pinch is the point of closest approach between the hot and cold composite curves as shown in Figure 19.166. It is also the point of zero energy flow in a network with minimum-utility usage. However, since any energy transfer across the pinch increases utility requirements, it is essential that for a minimum-utility design, there can be no energy flow across the pinch. This allows the heat exchanger network (HEN) design to be decomposed into two separate problems: the hot end problem above the pinch and the cold end problem below the pinch. The two problems can be solved separately and then combined to obtain a solution for the entire network. Due to the low driving force, exchangers operating near the pinch tend to be relatively large and expensive units. Therefore, identification of appropriate stream matches for exchangers operating at the pinch is a key aspect of the process integration methodology. Exchangers that operate at the pinch have a temperature difference equal to ΔT_{min} or at least at one end.

19.57 The Pinch Design Approach to Inventing a Network

In designing a heat exchanger network that uses the minimum utilities for maximum heat recovery, we consider the following steps [213]:

1. Select a ΔT_{min}.
2. Calculate the minimum hot and cold utilities to use based on this value for ΔT_{min}.
3. Using the grand composite curve, pick which utilities to use and their amounts.
4. If the problem has a pinch point in it (which will occur if step 2 discovers the need for both heating and cooling), divide the problem into two parts at the pinch. We shall design the two parts separately. Remember that the part above the pinch requires only hot utilities, and the part below only cold utilities.
5. Estimate the number of exchangers for each partition as N-1, where N is the number of streams in that part of the problem.
6. Invent a network using all insights available. All exchangers that exit at the pinch point will have the minimum driving force at that point. A small driving force for heat transfer implies a large area. The exchangers near the pinch will tend to be large. Therefore, bad design decisions near the pinch point will tend to be more costly. We should generally make design decisions in the vicinity of the pinch first.
7. Remove heat cycles if possible.

For the designer wishing to produce a design in achieving minimum utility targets, the following rules are:

- No heat transfer between process streams across the Pinch temperature.
- No external (utility) cooling above the Pinch temperature.
- No external (utility) heating below the Pinch temperature.

Violation of any of the above rules results in higher energy requirements than the theoretical minimum requirements and will adversely affect the energy efficiency.

An Excel spreadsheet program (Example 19.49.xlsx) developed from the book by Ian C. Kemp [212] was used for Example 19.49. Process integration using pinch analysis is further reviewed in Chapter 22 in Volume 4 of these series.

19.58 Appropriate Placement and Integration of Distillation Columns

Distillation columns absorb heat at the reboiler temperature and then reject heat at a lower (condenser) temperature. From process integration viewpoint, they operate on temperature and not heat [214]. Thus, columns may be integrated with the background process and/ or with each other to save on utility consumption. Thus, appropriate placement requires positioning a column either entirely above or below the pinch of the background process. If a column is placed above the pinch, then the condenser can reject heat into the process sink; for a column placed below the pinch, then the reboiler can accept heat from the process source. In both of these cases, there is a decrease in the hot and cold utility consumptions.

If a column is placed across the pinch, then the reboiler consumes heat from the process sink above the pinch and the condenser rejects heat into the process source below the pinch. The total energy requirements for the column and process together are no different from that for the separate systems. Thus, there are no energy savings from the integration. Additionally, there is a heat load limit for distillation columns, which requires that no infeasibility (i.e., negative heat flow) must occur in the cascade for the process over the entire temperatures spanned by the column.

For optimum performance, modifications related to reflux ratio, pressure, feed preheating/cooling, side condensers, and reboilers may be required. The grand composite curve (GCC) may be used to establish targets in terms of temperature and heat loads for the best combination of these options prior to column/process design. Shenoy [215] and Linnhoff *et al.* [214, 216, 217] provide details on column modifications.

19.59 Heat Integration of Distillation Columns: Summary

The appropriate placement of distillation columns when carrying out heat integration network is not across the pinch. Further, the grand composite curve as shown in Figure 19.168 can be used as a quantitative tool to assess integration opportunities. The scope for integrating conventional distillation columns into an overall process is restricted, and practical constraints often prevent integration of columns with the rest of the process. If the column cannot be integrated with the rest of the process or the potential for integration is limited by the heat flows elsewhere, then attention must be focused on the distillation operation and complex arrangement being considered.

Once the column is integrated, then the driving forces between the composite curves become smaller. This will enable the capital/energy trade-offs of the heat exchanger network to be adjusted accordingly.

Maximization of Crude Preheat

A good heat exchanger synthesis through a pinch analysis increases the preheat in the crude unit. The first thing required when carrying out a network synthesis or a pinch analysis is the heat source table as shown in Figure 19.178. The heat source table is developed from the results of simulation. A hot composite and cold composite, and grand composite curve (GCC) may be the next step after a simulation run, and a heat source table is constructed from the results of the simulation. The hot, cold, and grand composite curves (e.g., Figures 19.180 to 19.182) can be developed including the topper overhead heat as available heat.

In a crude distillation unit (CDU), cold streams are typically crude, demineralized water (DMW) for desalting, stabilizer feed, stabilizer reboiler heat, stripper reboiler heat, and so on. The cold stream flows along with their heat capacities are used for constructing the cold composite curve. The GCC is not extended beyond the achievable preheat (upper open jaw not shown in GCC).

Simulation plays the principal role in finalizing the heat source table—the hot streams, their terminal temperatures and their (mass flow rate) × (heat capacity) values. During simulation, the pump around (PA) duties can be increased, which is limited by the internal reflux (IR) in the zone above the PA return. Preference is given for an increase of heat duty in the high-temperature PA. The IR must remain above the minimum to maintain the efficiency of the trays between the topmost sidecut (kerosene or heavy naphtha) with top naphtha. Furthermore, a balance must be reached between the stages separating adjacent cuts and the IR in the section in order to achieve optimum design. An example of pinch analysis of the CDU is provided in chapter 22.

19.60 Common Installation Errors in Distillation Columns

Nag [225] provides problems caused by errors in installation and the correct arrangement. Strip-out (Figure 19.184) from the side stripper is introduced within the pump around trays (typically three trays). This arrangement has occurred in many columns and rectified during revamp by rerouting the strip-out above the PA return. If the strip-out is incorrectly introduced, then its vapor would condense and recirculate. The draw-off nozzle for PA trays should be flushed with the bottom of the pan; otherwise, a weep hole is required on the pan. The desired elevation of the stripper and seal in a two-pass tray column is shown in Figure 19.185.

Similar arrangements have showed in other columns, where the strip-out enters the tray above the two-pass tray. It can be seen that the liquid drawn from the tray is actually the liquid from the tray above. If the strip-out is returned to the location shown, then this would be introducing it below the tray and would recirculate the strip-out.

For two-pass trays, the strip-out will flow through half of the tray's active area as the central downcomer will not allow the vapor to bubble through the other half of the tray. The central and off-center downcomers are always provided with vapor-equalizing tunnels to equalize the pressure in each section constituted by the downcomers.

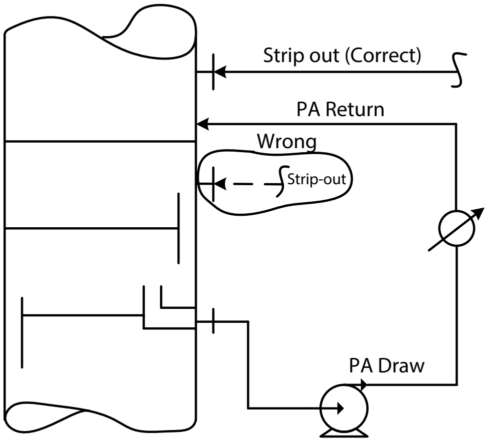


Figure 19.184 Strip-out wrongly entering between PA sections.

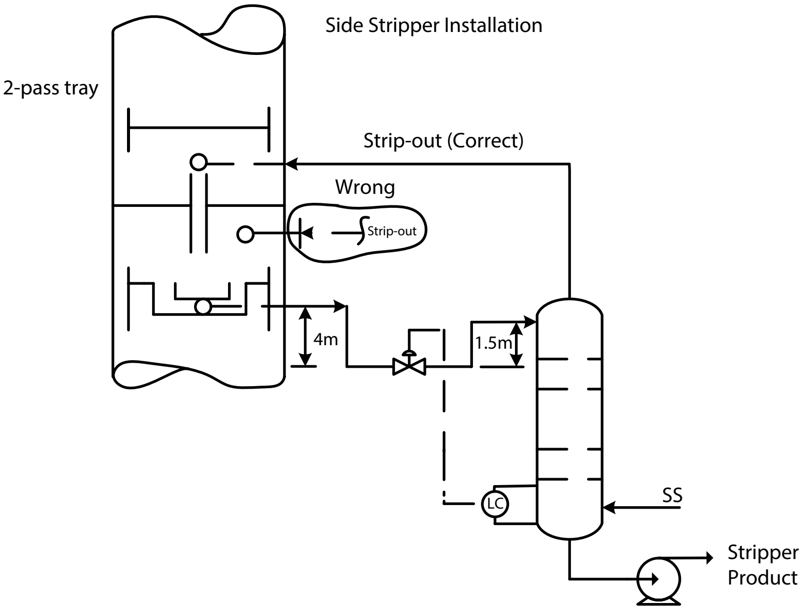


Figure 19.185 Desired elevation of stripper and seal in a two-pass tray column.

For side-stripper installation, the difference in elevation between the column draw tray and stripper top should be a minimum of 4 m as shown in Figure 19.185. This elevation difference provides the necessary head to overcome the pressure loss in the product draw line and draw-off control valve. A liquid seal of 1,500 mm is provided so that the strip-out vapor does not blow through the liquid draw line due to the pressure drop in the stripper overhead line joining the column, and the control valve loop provides the seal required.

Calculation of Nozzle Size [225]

Self-Venting Withdrawal Nozzles and Lines

Liquid outlet line from the column should meet the self-venting criteria. The liquid draw from the distillation column is partly aerated. The size of the self-venting line should provide enough free area for the released vapor to return to the column; otherwise, it could cause cavitation in the pump.

Appropriate self-venting nozzle size is estimated based on the velocity in the nozzle, which ranges from 0.6 to 1.2 m/s for side draw-off. The table below shows the considered velocity in the nozzle for the range of flow of liquid draw from the column.

Liquid Flow From the Column m^3/h	**Considered Velocity in the Nozzle m/s**
300	0.6–0.7
300–500	0.9
500–1,000	0.9–1.0
1,000 and higher	1.0–1.2

Some designers use a velocity of 0.9 m/s at all ranges of flow, but a conservative sizing may be better. Some generalized correlations are suggested by experts and practitioners. The size reduction (swage down) is not allowed before a 3-m vertical drop of the withdrawal line from the draw-off nozzle. The draw-off line may have bends/elbows after a minimum vertical descent of 3 m. Bends may have a long radius elbow.

Column Inlet Nozzle

The diameter of the column inlet nozzle (the flash zone of the vacuum distillation column) is calculated with a maximum gas velocity of 0.6–0.7 times the velocity of sound in the media, or as a maximum of 350 ft/s (107 m/s).

The velocity of sound is calculated $= 31.6\left(\frac{kP}{\rho}\right)^{0.5}$

where

k = ratio of specific heat capacities (C_p/C_v) of the vapor.
P = absolute pressure, kPa.
ρ = density of the gas, kg/m^3.

Alternatively, it can be calculated by the expression:

$$V_s = 6{,}160\left(\frac{kT}{1.1M_w}\right)^{0.5}, \frac{m}{s}$$

where

V_s = velocity of sound m/s
T = temperature, K (= °C + 273.15)
M_w = molecular weight

Often for vacuum transfer lines or nozzles, the diameter thus calculated is divided by 0.98 to account for the flow of liquid.

Typical Residence Time in Vessels—Tower Bottom

1. The preferred time for column bottom sump being pumped to furnace is 10 min. This amount can be less if the nozzle is self-venting. In addition, 3–5 min has been applied in several applications [225].
2. Column bottom pumped to storage is 2 min.
3. Reflux to column is 5 min.

4. Column bottom pumped to reboiler furnace is 5 min.
5. Residence time for withdrawal from chimney trays is normally less than 1 min. in order to contain the height of chimney. External vessels can be provided if higher time is desired for the pump.

Residence time is calculated based on the volume of the vessel/tower between the low-level alarm point and a high-level alarm point. The low-level alarm point is normally at 200–300 mm above the bottom tangent line of the towers. If the residence time is known, then the high-level alarm point can be determined using the residence time.

Steps in the Design of a Fractionating/Distillation Column

The following steps are involved in the design of a fractionating/distillation column.

1. Specify the degree of separation or set the product specification/purities.
2. Select the operating conditions (pressure of the column, the reflux drum pressure, etc.)
3. Determine the stages and reflux requirements.
4. Select the type of contacting device (plates or packing).
5. Size the column based on diameter, number of actual trays, etc.
6. Design the column internals (plates, distributors, packing supports, etc.)

Column Top Temperature

The top column temperature is the dew point of the overhead vapor at the partial pressure of the hydrocarbons. The hydrocarbon partial pressure is estimated from the hydrocarbon vapor phase mole fraction at the column top multiplied by the total pressure at the column top.

Manual methods are no longer used for design or rating because the heat and mass balances performed to derive the vapor-liquid loading in each section are very tedious and cumbersome and not practical for detailed design. Simulators are now being employed for design and rating and examples are provided in this chapter as well as volume 1 of these volume series.

Crude Column Simulation and Design

Inputs Required

1. Crude true boiling point (TBP) data (essential).
2. Crude density/American Petroleum Institute (°API) gravity, (essential).

$$(^\circ\mathrm{API}) = \frac{141.5}{\mathrm{Sp\,Gr}} - 131.5$$

3. Crude oil molecular weight (Optional).
4. Viscosity (optional).

Specifications Required

1. Column pressure/reflux drum pressure.
2. Fixed product draw rates or distillation points.
3. PA duties and return temperature or temperature differentials.
4. Column top temperature or overhead product rate or distillation end point of overhead product.

Typical Number of Stages in the Sections of a Crude Unit

1. Between top naphtha and superior kerosene (SK): 5–8 stages. (It is a function of naphtha end point and SK flash.)
2. Between SK and diesel: 4 to 5 stages. (It is a function of the specified gas or overlap between SK and diesel.)
3. Wash section (above flash zone and lowermost draw): 2–3 stages.
4. Bottom stripping section: 2–3 stages.

Tray Efficiencies and System Factors to be Considered for Design

Pre-topper

Efficiency:	65%–70%
System factor:	0.8–0.85

Topper

Efficiency above wash zone:	60%–65%
Efficiency of wash zone:	30%–40%
Efficiency of stripping zone:	30%
System factor:	0.85–0.9

Typical Simulation Inputs

Thermodynamics (K-value):	Grayson-Streed (GS)/Soave-Redlich-Kwong (SRK)
Transport property:	PETRO/SIMCI heavy oil correlation. The viscosity of crude and heavy fractions may be estimated from a mid-percent viscosity curve given along with crude TBP and can be entered in the simulation.
Density:	°API
Kerosene PA differential temperature (DT)	55°C–60°C
LGO PA (DT):	55°C–60°C
HGO PA (DT):	55°C–60°C
Typical reflux ratio:	1.5–1.8

DT is the differential temperature between pump-around supply and return.

Pump-Around Reflux Heat Duties and Differential Temperatures

Maximizing PA duties are essential for maximum energy efficiency. This is limited by gap or overlap specifications between adjacent products and a minimum IR between trays. The IR in a section controls the separation between the cuts with a fixed number of trays in the section.

A differential temperature between PA supply and return temperature is typically specified as 55°C–60°C. A larger temperature difference would lower the flow of PA refluxes but also reduce the differential temperature across crude preheat exchangers, which can prevent heat for the crude at higher temperature ranges.

Sometimes, when the pump-around section of the column is overloaded due to high flow in the section, the DT of the PA is increased to reduce the PA flow, and part of the PA heat is passed to the lower temperature section of the crude preheat train and may result in an inefficient design.

Location of Pump-Around Refluxes

Figure 19.186 shows different arrangements of PA refluxes, as these can cool the vapors and condense them to be drawn as products from below the PA return. A PA intended to condense a product can be drawn from the same tray (Type B in Figure 19.186), then cooled and returned three trays above. Some amount of IR joins the PA return from the top and condenses the cuts below.

A PA can also be located above the draw tray below the upper sidecut. In this case, the PA will generate high IR to the flow down the stages to the draw tray (Type A in Figure 19.186). This arrangement helps create a better separation of the adjacent cuts. Separation is a function of the IR multiplied by the stages in the section. Thus, separation between the cut and the cut above improves in this arrangement. However, if the PA is placed at an upper location, the PA draw temperature reduces and the quality of PA heat becomes inferior. This may affect the heat recovery and thermal efficiency. A combination of both types can also be practiced, if adequate stages are available in the section.

A PA can also be configured as a conservative intercooler (Type C in Figure 19.186). In this situation, a part of the PA can be cooled and then returned below the draw tray to cool and condense the products located below.

Typical Stabilizer Design Parameters

Typical number of actual trays in the Stabilizer:	44–50
Reflux ratio:	2–2.5
Efficiency of trays	65%–70%
System factor:	0.9
Thermodynamics: VLE (K-value):	Peng-Robinson (PR), SRK, GS
Transport property:	PETRO
Density:	°API

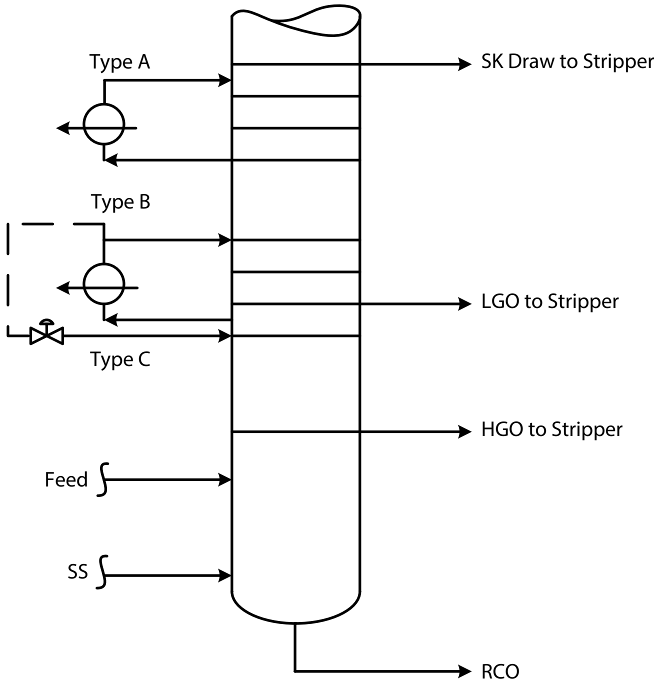


Figure 19.186 Different arrangements of pump-around refluxes.

Typical Naphtha Splitter Design Parameters

Typical number of actual trays in The splitter:	40–60
Reflux ratio:	1.5–3.5
Efficiency of trays	65%–70%
Thermodynamics: VLE (K-value):	GS
Transport property:	PETRO
Density:	°API

Water Dew Point Calculation of a Crude Column

Overhead product is D, kg/h
Reflux rate is R kg/h
Total overhead vapor, V = (D + R), kg/h
kg moles HC vapor at the top (M_{HC}) = (D + R)/molecular weight
Flow of stripping steam = S1 kg/h (including side strippers)
Molecular weight of top naphtha ranges from 90 to 100.
Water along with crude = S2 kg/h (from the desalter)
1% wt of water along with reflux = (1 × R)/100 = S3 kg/h
Total water at the top of the column (S kg/h) = S1 + S2 + S3
Moles of water at the top of the column $\left(M_{H_2O}\right) = \frac{S}{18}$ kmoles/h
Mole fraction of water = $y_{H_2O} = \frac{M_{H_2O}}{\left(M_{HC} + M_{H_2O}\right)}$
Partial pressure of water at the top of the column $\left(p_{H_2O}\right) = y_{H_2O} \text{ x } P_{top}$
P_{top} = top pressure, kg/cm^2, abs

At the dew point, the partial pressure of water at the top is equal to the vapor pressure or saturation pressure of water at that temperature. The steam table is used and the water saturation temperature corresponding to p_{H_2O} can be found. The saturation temperature is the water dew point at the top of the column. The top temperature should be kept above the water dew point by around 6°C–10°C to avoid water condensation and corrosion at the upper section of the column.

If there is a top PA in the column, then it should return to the column top with a coalescer for water removal. Otherwise, heavy corrosion may result at the uppermost part of the column. Usually, the top section is lined with Monel to avoid corrosion.

Advanced Distillation Technologies

Cyclic Distillation

Conventional distillation processes are energy intensive and inefficient and would require optimization to reduce energy (i.e., pinch technology) and capital costs to improve economic efficiency. Novel distillation concepts based on process intensification (PI) principles can provide major advantages in terms of significantly lower energy use and reduced capital costs while improving environmental efficiency.

The PI concept has received attention in recent times as the objective is to design substantially smaller plants while improving their operational safety, environmental performance, and energy efficiency. The Rapid Advancement in Process Intensification Deployment (RAPID) Institute, supported by the U.S. Department of Energy has promoted major advances in energy efficiency and productivity in the process industries. With PI, distillation has been enhanced such as HiGee distillation, dividing-wall column, heat-integrated distillation, reactive distillation and membrane distillation (MD) [332–335].

Another technology is cyclic distillation in fluid separations due to a different way of contacting the liquid and vapor phases. In contrast to classic operation, cyclic distillation uses separate phase movement (SPM) that can be achieved with specific internals and a periodic operation mode. Cyclic distillation replaces the old distillation columns with new ones using a cyclic operation mode. One operating cycle consists of two key parts: a vapor flow period (when the thrust of rising vapor prevents liquid down flow) followed by a liquid flow period (when the liquid flows down the column, dropping by gravity, first to a lock chamber and then moving to the tray below).

Figure 19.187 illustrates the operating cycle. A cyclic distillation resembles a regular column. However, the cross-section view inside a cyclic distillation column shows the absence of downcomers and the presence of different internals that allow an efficient separate-phase movement in practical operation. The process is based on the sequential supply of vapor and liquid to the column. During the vapor period, rising vapor up the column prevents liquid down flow by closing a set of valves on each tray (Figure 19.187a). Vapor runs through the liquid via perforations on the bottom of the tray. The liquid is assumed to be perfectly mixed without a temperature or concentration gradient, and all the liquid on the tray has the same residence time and a speed of movement equal to zero. During the liquid period, the valves open and liquid flows into a sluice chamber located under each tray (Figure 19.187b). As the vapor period starts again, the valves close and liquid moves from the sluice chamber to the tray below (Figure 19.187c).

Theory

The algorithms for cyclic distillation columns include analytical and numerical methods of increased complexity, and Figure 19.188 shows the modeling of a classic theoretical stage and the operating lines in case of classic vs. cyclic distillation—where L and G are the liquid and gas (vapor) flow rates, H is the liquid holdup on the tray (mol), τ is the vapor period (sec), and x and y are the liquid and vapor compositions, respectively.

Figure 19.189 shows the operating lines for binary separations in the classic distillation vs. cyclic distillation. The operating lines are the basis of designing distillation columns, and a distinction arises from the theoretical stage concepts. This difference is shown by the effectiveness of the minimum number of theoretical plates required for the separation at total/infinite reflux. The operating lines provide the number of stages, which is much lower as the case of cyclic distillation, due to the larger driving force of the process. The operating line is an actual mirror image of the equilibrium line relative to the diagonal.

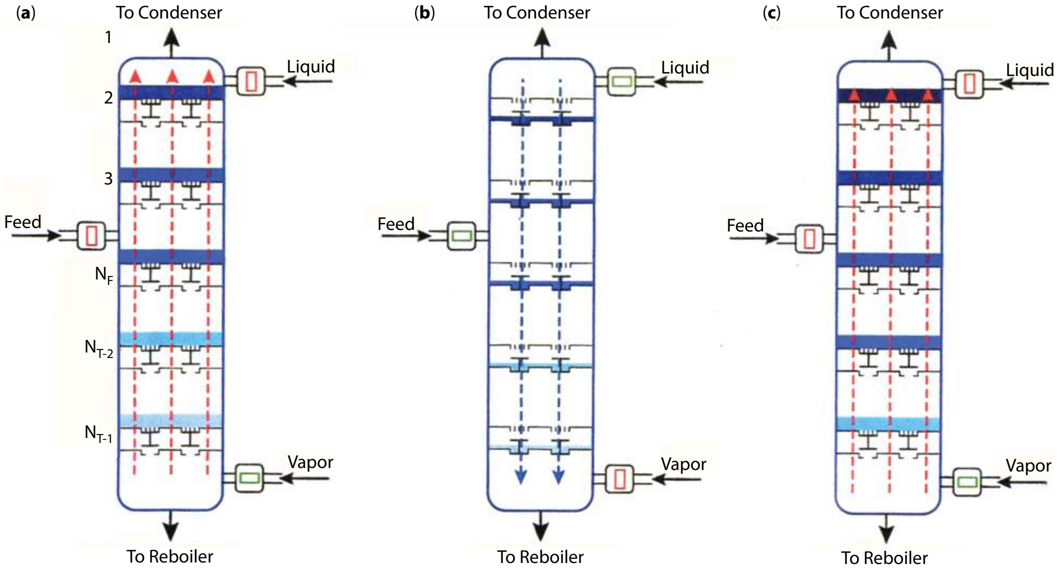


Figure 19.187 Cyclic distillation (Source: Stankiewicz., A., The Principles and Domains of Process Intensification, CEP, pp 23–28, March 2020 [232]).

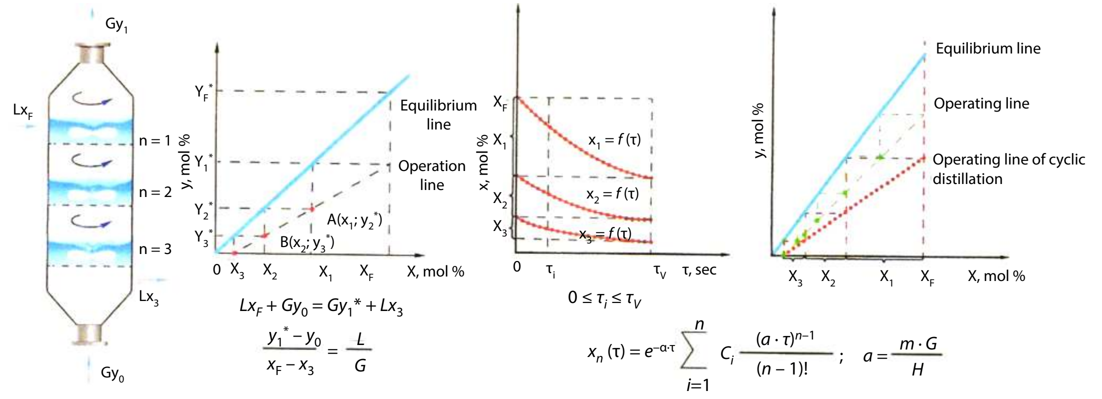


Figure 19.188 Conventional theoretical stage and operating lines (left); operating lines for cyclic mode (right) (Source: Kiss, A., *et al.*, [236]).

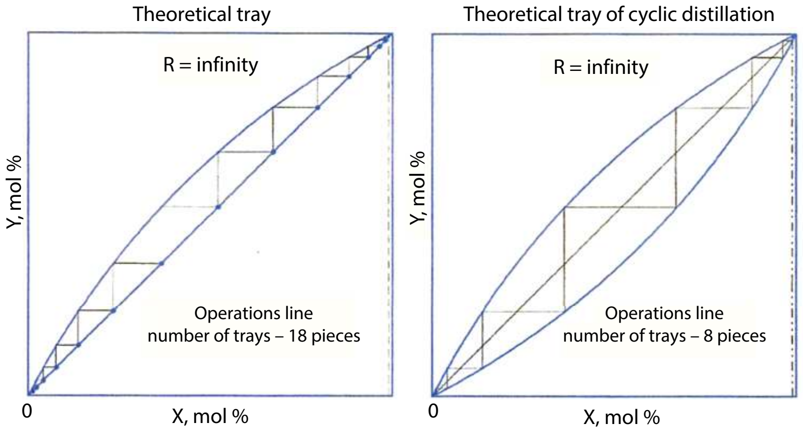


Figure 19.189 Operating lines for the separation of a binary mixture in case of classic (left) and cyclic distillation (right) (Source: Kiss, A., *et al.*, [236]).

An integrated process design and control approach was proposed based upon the application of simple graphic-design methods known from conventional distillation columns. This approach has shown that operating a cyclic distillation column at the largest driving force results in an optimum design in terms of controllability and operability, which is less sensitive to disturbances in the feed and has the inherent ability to reject disturbances [236].

Techno-Economic Advantages

The economic advantages derived from cyclic distillation are as follows [236]:

1. 20%–50% lower investment cost (due to lower column height, smaller column diameter, smaller area of the heat exchangers, less steel construction, and less space used).
2. 20%–35% lower hot utility usage (due to high mass transfer efficiency and reduction of the reflux rate).
3. 20%–35% lower cold utility usage (due to lower reflux rate).
4. Improvement of product quality (due to high mass transfer efficiency and higher concentration of key product).
5. Increase of product yield (due to high mass transfer efficiency and more concentrated impurities or other fractions).
6. Enhanced process sustainability (due to less GHG emissions due to lower energy usage).

The technology advantages of cyclic distillation are as follows [236]:

1. Mass transfer efficiency of one cyclic distillation tray is equal to three classic distillation trays.
2. Reduction of the residence time of liquid in the column, and the uniform arrangement of liquid on the tray.
3. Ability to control the amount of liquid on the tray and the reaction time (in the case of reactive distillation).
4. Any geometric configuration of the trays, which allows the possibility to build dividing-wall columns with trays.
5. The separation efficiency does not depend on the column diameter, which allows easy industrial scale-up.
6. Placement of any type of packing between the trays further increases the mass transfer efficiency.
7. The pressure drop in the column does not depend on the liquid load in the column since the amount of liquid on the trays is constant and only the frequency of cycles is changed (range: 1 to 30 m^3/m^2 h liquid load).
8. The vapor velocity in the column typically ranges from 0.2 to 20 m/s, as it depends on the pressure in the column.
9. The operation remains stable and efficient in case of changed concentrations of the key components.

Disadvantages

1. Cyclic distillation is characterized by limitations when considering the technology for various applications.
2. It is difficult to apply to vacuum systems due to the pressure drop on the trays.
3. The performance enhancement critically depends on the separation of the vapor and liquid periods.

Applications

Cyclic distillation has been employed in industrial applications due to the availability of design and control methods and the introduction of special trays that allow better control of phase movement. New internals have provided the industrial breakthrough of cyclic distillation technology. The cyclic operation uses reliable stop valves (butterfly valves with pneumatic drive) from manufacturers, which can withstand up to 20 MM cycles (e.g., the first industrial columns operated for more than 10 years).

Furthermore, more efficient contact devices (e.g., trays with lock chambers allowing the liquid to follow the path of the tray to lock chamber to the tray below, see Figure 19.190), and Figure 19.191 shows a full column. Practical implementation at pilot and large scale shows the following:

- High tray efficiencies (140%–200% Murphree efficiency) translate into reduced equipment cost.
- Higher throughput and equipment productivity than conventional distillation.

Figure 19.190 Trays suited with sluice chambers, especially designed for cyclic distillation columns (left). Top view of an installed cyclic distillation tray column with a diameter of 1.7 m (right) (Source: Kiss, A., *et al.* [236]).

Figure 19.191 Cyclic distillation (dividing wall) column prepared for transportation to a chemical site (Source: Kiss, A., *et al.* [236]).

- Reduced energy requirements (20%–35% savings) translate into lower operating costs.
- Increased product quality due to higher separation efficiency.

Although cyclic distillation is mostly employed in the food industry, however, it is being employed in areas of high-potential use, such as oil refining, chemicals, petrochemicals, pharmaceuticals, biofuels, and so on. The technology has been shown to improve fluid separations for processing hydrocarbons and their derivatives, providing key benefits such as follows:

- Lower energy requirements (by 20%–35%).
- Reduced capital costs (by 20%–50%).
- Increased column throughput.
- Better separation performance (with up to three times higher efficiency than classic distillation trays).
- Provides more degrees of freedom contributing to good process control and simple operation.

This technology requires enhanced inputs by chemical process industry and hydrocarbon process industry to realize its potential for the enormous benefits and advantages over the classic distillation technique.

Membrane Processes

Membrane technology has emerged as growing technology in separation processes and has been increasingly applied in waste-water treatment, desalination, gas separation purification, and product specification. Membrane distillation (MD) is a thermally driven separation process in which the separation is driven by phase change. A hydrophobic membrane presents a barrier for the liquid phase, allowing the vapor phase (e.g., water vapor) to pass through the pores of the membrane. The driving force of the process is a partial vapor pressure difference commonly initiated by a temperature difference.

Most processes that use a membrane to separate materials rely on the static pressure difference as the driving force between the two bounding surfaces [e.g., reverse osmosis (RO)] or a difference in concentration (dialysis), or an electric field. The selectivity of a membrane can be due to the relation with the pore size to the size of the substance being retained, or its diffusion coefficient, or its electrical polarity. Membranes used for MD inhibit passage of liquid water while allowing permeability for free water molecules and thus for water vapor. These membranes are made of hydrophobic synthetic material (e.g., PTFE, PVDF, or PP) and offer pores with a standard diameter between 0.05 and 0.5 μm. As the water has strong dipole characteristics, while the membrane fabric is non-polar, the membrane material is not wetted by the liquid. Even though the pores are considerably larger than the molecules, the high-water surface tension prevents the liquid phase from entering the pores. Figure 19.192 shows the principle of membrane separation process.

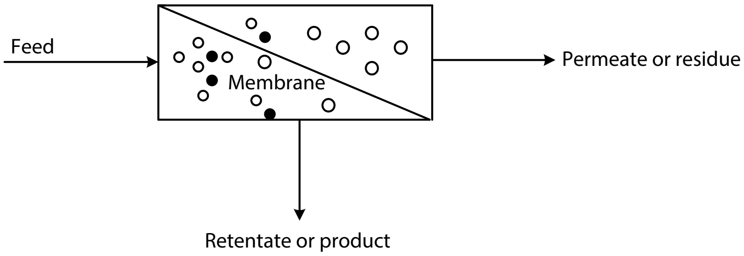
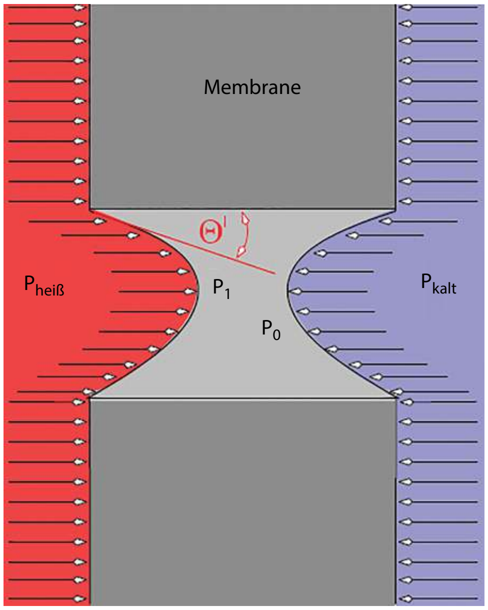


Figure 19.192 Principle of the membrane separation process.

The driving force which delivers the vapor through the membrane in order to collect it on the permeate side as product water, is the partial pressure water vapor pressure difference between the two bounding surfaces. This partial pressure difference is the result of a temperature difference between the two bounding surfaces. The membrane is charged with a hot feed flow on one side, and a cooled permeated flow on the other side, as the temperature difference is 5–20 K. This ensures a partial pressure difference that the vapor developing at the membrane surface follows

Capillary depression of water on a hydophobic membrane



Temperature and pressure profile through the membrane considering temperature polarisation

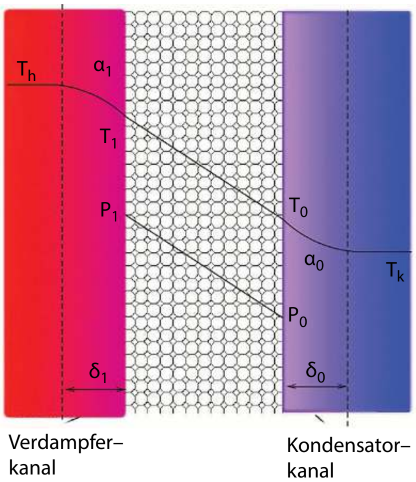


Figure 19.193 Temperature and pressure profiles through the membrane (Source: www.wikipedia.com).

Table 19.62 Membrane material.

Modified natural product	Cellulose acetate, cellulose autobutyrate, cellulose nitrate
Synthetic products	Polyamide, polybenzimidazole, polysulfone, vinyl polymers, polyfuran, polycarbonate, polyethylene, polypropylene, polyvinyl acetate, polyacronitrile, polytetrafluoroethylene
Miscellaneous	Polyelectrolyte complex, porous glass, graphite oxide, ZrO_2, polyacrylic acid, ZrO_2-carbon, etc.

Table 19.63 Gas membrane application area.

Common gas separation	**Application**
O_2/N_2	Oxygen enrichment, inert gas generation
H_2/Hydrocarbons	Refinery hydrogen recovery
H_2/CO	Syngas ratio adjustment
H_2/N_2	Ammonia purge gas
CO_2/Hydrocarbons	Acid gas treatment, landfill gas upgrading
H_2O/Hydrocarbons	Natural gas dehydration
H_2S/Hydrocarbons	Sour gas treating
He/N_2	Helium recovery
Hydrocarbon/Air	Hydrocarbon recovery, pollution control
H_2O/Air	Air dehumidification

the pressure drop, permeating through the pores and condensing on the cooler side. Figure 19.193 shows the temperature and pressure profiles through the membrane considering the temperature polarization. Table 19.62 shows commercially viable membranes for CO_2 removal are polymer based cellulose acetate, polyamides, polysulfone, polycarbonate, and polyetherimide. Gas separation membranes are manufactured in one of two forms: flat sheet and hollow fiber (Table 19.63).

The four MD techniques that differ mainly by the arrangement of their distillate channel or the manner in which this channel is operated are as follows:

- Direct contact membrane distillation (DCMD)
- Air gap membrane distillation (AGMD)
- Vacuum membrane distillation (VMD)
- Sweeping gas membrane distillation (SWGMD)
- Vacuum multi-effect membrane distillation (VMEMD)
- Permeate gap membrane distillation (PGMD)

Direct Contact Membrane Distillation (DCMD)

In DCMD, both sides of the membrane are charged with liquid-hot feed water on the evaporator side and cooled to permeate on the permeate side. The condensation from the vapor passing through the membrane happens directly inside the liquid phase at the membrane boundary surface. Since the membrane is the only barrier blocking the mass transport, relatively high surface related permeate flows can be achieved with DCMD [227]. The disadvantage is the high sensible heat loss as the insulating properties of the single membrane layer are low. However, a high-heat loss between the evaporator and condenser is the result of the single membrane layer. Although this heat loss is not available in the distillation process thus reducing the efficiency [228]. In the DCMD, the cooling across the membrane is provided by the permeate flow rather than the feed preheating. Therefore, an external heat exchanger is required to recover the heat from the permeate, and the high-flow rate to the feed must be carefully optimized.

Air Gap Membrane Distillation (AGMD)

In AGMD, the evaporator channel resembles that in DCMD; however, the permeate gap lies between the membrane and a cooled wall that is filled with air. The vapor passing through the membrane must additionally overcome this air gap before condensing on the cooler surface. The advantage with this technique is the high thermal insulation toward the condenser channel, thus reducing the heat conduction losses. Furthermore, the volatile substances with a low surface tension such as alcohol or other solvents can be separated from diluted solutions. This is because there is no contact between the liquid permeate and the membrane with the AGMD. The disadvantage is that the air gap represents an additional barrier for mass transfer, reducing the surface related to the permeate output compared to DCMD [229]. The AGMD is advantageous compared to alternatives at high salinity and variations can include hydrophobic condensing surfaces for improved flux and energy efficiency. Its features in the design include gap thickness, condensing surface hydrophobicity, gap spacer design, and tilt angle [230]. Figure 19.194 shows droplet condensation regimes in the AGMD.

Sweeping Gas Membrane Distillation (SWGMD)

The SWGMD also referred to as air stripping uses a channel configuration with an empty gas on the permeate side. This configuration is identical with AGMD. Condensation of the vapor takes place outside the membrane module in an external condenser. As with AGMD, volatile substances with a low surface tension can be separated during the process. The advantage of SWGMD over AGMD is the significant reduction of the barrier to the mass transport through forced flow. A higher surface related mass flows can be achieved than with AGMD. The disadvantage of SWGMD is caused by the gas component and therefore, the higher total mass flow, is the necessity of a higher condenser capacity. When using smaller gas mass flows, there is a risk of the gas heating itself at the hot membrane surface, thus, by reducing the vapor pressure difference invariably reduces the driven force. One solution to the problem for SWGMD and AGMD is employing a cooled walling for the permeate channel, and maintaining the temperature by flushing it with gas.

Vacuum Membrane Distillation (VMD)

VMD contains an air gap channel configuration. Once it has passed through the membrane, the vapor is sucked out of the permeate channel and condenses outside the module as with the SWGMD. VCMD and SWGMD can be applied for the separation of volatile substances from a watery solution or for the generation of pure water from

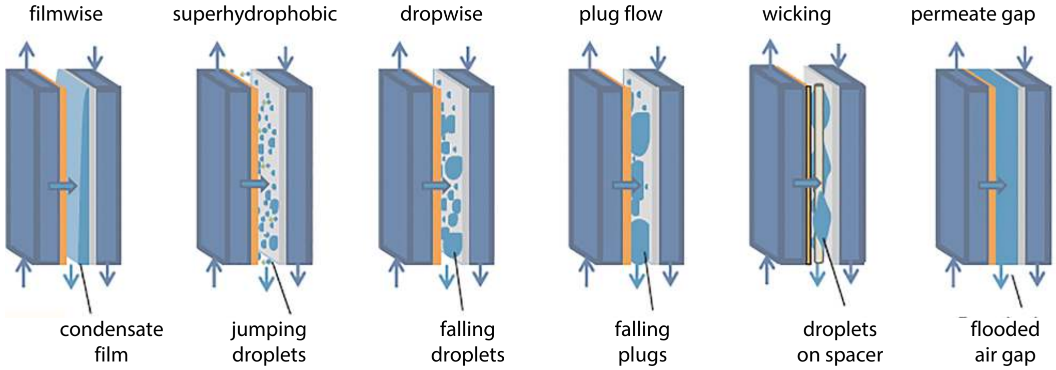


Figure 19.194 Droplet condensation regimes in AGMD.

concentrated salt water. The advantage of this method is that the undissolved inert gases blocking the membrane pores are sucked out by the vacuum, leaving a larger effective membrane surfaces active. Furthermore, a reduction of the boiling point results in a comparable amount of product at lower overall temperatures and lower temperature differences through the membrane. A lower required temperature difference leaves a lower total and specific thermal energy demand. The disadvantage of VCMD is the generation of a vacuum, which must be adjusted to the salt water temperature and this requires complex technical equipment. The electrical energy demand is higher as with DCMD and AGMD; an additional problem is the increase of the pH value due to the removal of CO_2 from the feed water. An efficient VDMD requires having a multistage configuration of the unit.

Permeate Gap Membrane Distillation (PGMD)

The PGMD complete channel configuration consists of a condenser channel with an inlet and outlet and an evaporator channel with an inlet and outlet. These two channels are separated by the hydrophobic, micro porous membrane. For cooling, the condenser channel is flooded with fresh water and the evaporator with salty feed water. The coolant enters the condenser channel at a temperature of 20°C. After passing through the membrane, the vapor condenses in the cooling water, releasing its latent heat and resulting in a temperature increase in the coolant. Sensible heat conduction also heats the cooling water through the surface of the membrane. Due to the mass transport through the membrane, the mass flow in the evaporator decreases while the condenser channel increases by the same amount. The mass flow of the preheat coolant leaves the condenser channel at a temperature of about 72°C and enters a heat exchanger, thus preheating the feed water. This feed water is then delivered to a further heat source and finally enters the evaporator channel of the MD module at a temperature of 80°C. The evaporation process extracts latent heat from the feed flow, which cools down the feed increasingly in flow direction. An additional heat reduction occurs due to sensible heat passing through the membrane. The cooled feed water leaves the evaporator channel at approximately 28°C. The total temperature differences between the condenser inlet and the evaporator outlet and the condenser outlet and the evaporator inlet are about equal. In the PGMD module, the permeate channel is separated from the condenser channel by a condensation surface. This allows the direct use of a salt water feed as a coolant, since it does not come into contact with the permeate. In this aspect, the cooling or feed water entering the condenser channel at a temperature T1 can also be used to cool the permeate. Condensation of vapor takes place inside the liquid permeate. The preheated feed water that was used to cool the condenser can be conducted directly to a heat source for final heating, after leaving the condenser at a temperature T2. After it has reached temperature T3, it is guided into the evaporator. The permeate is extracted at temperature T5, and the cooled brine is discharged at temperature T4.

The advantage of PGMD over DCMD is the direct use of feed water as cooling liquid inside the module and therefore the necessity of only one heat exchanger to heat the feed before entering the evaporator. In this case, the heat conduction losses are reduced and expensive components can be cut. A further advantage is the separation of permeate from the coolant. Therefore, the permeate does not have to be extracted later in the process and the coolant's mass flow in the condenser channel remains constant. However, the low flow velocity of the permeate in the gap is a disadvantage of this configuration, as it leads to a poor heat conduction from the membrane surface to the condenser walling. High temperatures on the permeate side of the membrane bounding surface are the result of this effect, which lowers the vapor pressure difference and thus the driving force of the process.

Vacuum Multi-Effect Membrane Distillation (VMEMD)

The VMEMD module consists of a steam raiser, evaporation-condensation stages, and a condenser. Each stage recovers the heat of condensation, providing a multiple-effect design. The distillate is produced in each evaporation-condensation stage and in the condenser (Figure 19.195).

Evaporation-condensation stages: Stages are composed of alternative hydrophobic membrane and foil [polypropylene (PP)] frames. Feed (e.g., seawater) is introduced into stage 1 of the module; it flows serially through the evaporation-condensation stages, and it exists at the end of the last stage as brine.

Stage 1: Steam from the evaporator condenses on a PP foil at the pressure level P1 and corresponding temperature T1. The combination of a foil and a hydrophobic membrane creates a channel for the feed, where the feed is heated

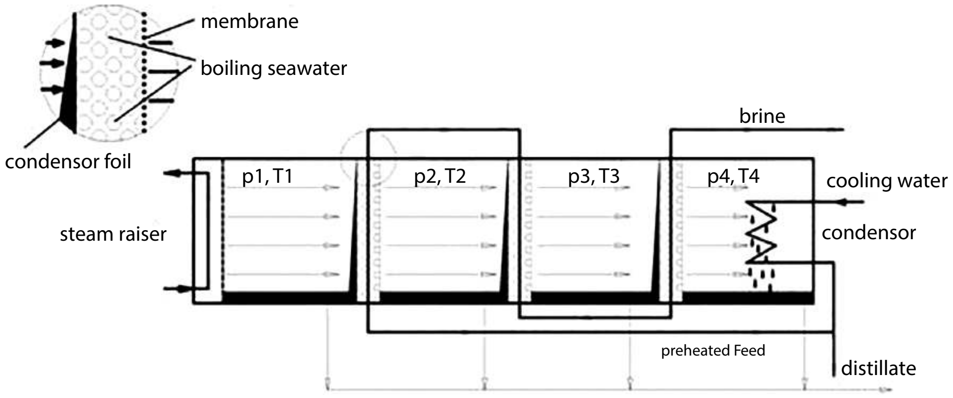


Figure 19.195 Diagram of memsys V-MEMD process (Source: www.wikipedia.com).

by the heat of condensation of the vapor from the steam raiser. The feed then evaporates under the negative pressure P2. The vacuum is always applied to the permeate side of the membranes.

Stages 2, 3, 4, …, n: This process is replicated in further stages, and each stage is at a lower pressure and temperature.

Condenser: The vapor produced in the final evaporation-condensation stage is condensed in the condenser, using the coolant flow (e.g., seawater).

Distillate production: The condensed distillate is transported via the bottom of each stage by pressure difference between the stages.

Design of memsys module: Inside each memsys frame, and between the frames, channels are created. Foil frames are the distillate channels. Membrane frames are the vapor channels. Between the foil and membrane frames, feed channels are created. The vapor enters the stage and flows into parallel foil frames. The only option for the vapor entering the foil frames is to condense, that is vapor enters a dead-end foil frame. Although it is called a dead-end frame, it does contain a small channel to remove the non-condensable gases and to apply the vacuum.

The hydrophobic membrane (or PP foils) is welded at both sides of the memsys frame. This frame is designed to combine and distribute vapor, feed, non-condensable gas, and distillate flows.

Figure 19.196 The hydrophobic membrane (Source: www.wikipedia.com).

Different numbers of memsys frame are vibration welded as memsys module (e.g., steam raiser, membrane stage, and condenser). GOR and capacity of memsys module can be easily modified depending on the application or customer's needs.

Figure 19.197 Different numbers of memsys frame (www.wikipedia.com).

The condensed vapor flows into a distillate channel. The heat of condensation is transported through the foil and is immediately converting into evaporation energy, generating new vapor in the seawater feed channel and is collected in a main vapor channel. The vapor leaves the stage via this channel and enters the next stage. Memsys had developed a highly automated production line for the modules and could be extended. As the memsys process works at modest low temperature (<90°C) and moderate negative pressure, all module components are made of PP. This eliminates corrosion and scaling and allows large-scale cost efficient production. Figures 19.196 and 19.197 show different numbers of memsys frame and vibration welded as memsys module, respectively.

Process Safety Incidents

The following cases studies provide root causes, key lessons/findings, and recommendations with reference to the pertinent chapters in these volumes' series (Volumes 1–4) for detailed explanations in designs and various process safety measures.

Case Study 1

BP Refinery Explosion, Texas City, USA, 2005

An explosion and fires occurred within the Isomerization unit (ISOM)[†] of BP's Texas City Refinery during a startup after a turnaround on March 23, 2005. Fifteen contractors were killed and over 170 people sustained injury, alarmed the community, and resulted in financial losses exceeding $1.5 billion. The incident occurred during the startup of an isomerization unit when a raffinate splitter tower[††] was overfilled; pressure relief devices opened, resulting in a flammable liquid geyser from a blowdown stack that was not equipped with a flare. The release of flammables led to an explosion and fire. All the fatalities occurred in or near office trailers located close to the blowdown drum.

Process Description of the Isomerization (ISOM) Unit

Figure 19.198 shows the process flow diagram of the isomerization (ISOM) unit. The raffinate splitter section took raffinate—a non-aromatic, primarily straight-chain hydrocarbon mixture from the aromatic recovery unit (ARU)

[†] The refining isomerization process converts straight chain normal pentane and normal hexane streams to the higher octane branched hydrocarbons isopentane and isohexane that are used for gasoline blending.

[††] The raffinate splitter is a distillation tower that takes raffinate, a non-aromatic, primary straight-chain hydrocarbon mixture and separates it into light and heavy components.

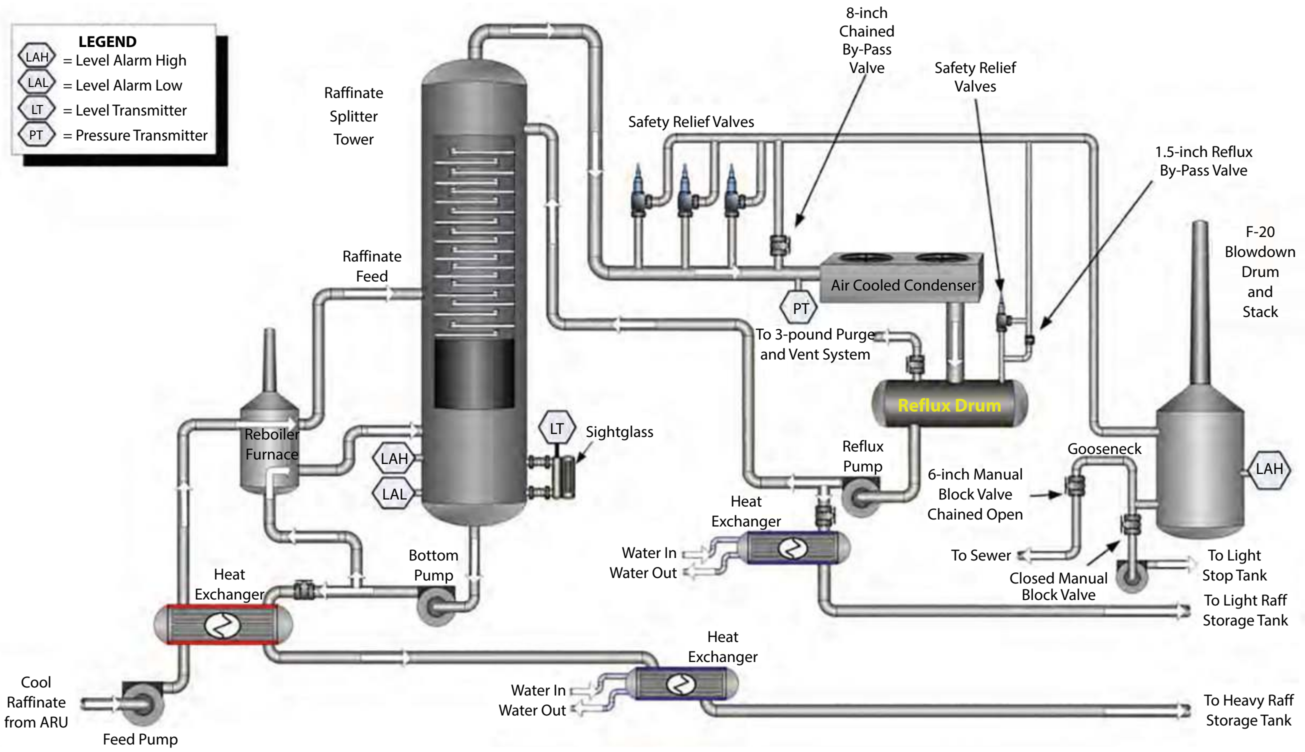


Figure 19.198 Process flow diagram of the Raffinate Column of the ISOM and the blowdown drum (Source: www.csb.gov).

and separated it into light and heavy components. About 40% of the raffinate feed was recovered as light raffinate (primarily pentane/hexane, C_5H_{12}/C_6H_{14}). The remaining raffinate feed was recovered as heavy raffinate, which was used as a chemical feedstock, jet fuel or blended into unleaded gasoline. The raffinate splitter unit could process up to 45,000 bpd of raffinate feed. Figure 19.198 consisted of a feed surge drum, a distillation tower; a furnace with two heating sections, one used as a reboiler for heating the bottoms of the tower and the other preheating the feed; air-cooled fin fan condensers and an overhead reflux drum; various pumps and heat exchangers. The distillation tower was a vertical column with an inside diameter of 12.5 ft (3.8 m) and height of 170 ft (52 m) with an approximate liquid-full volume of 154,800 gallons (586,100 liters). The tower was fitted with 70 distillation trays[§] that separate the light from the heavy raffinate (Figure 19.198).

Liquid raffinate feed was pumped into the raffinate splitter tower near the tower's midpoint. An automatic flow control valve adjusted the feed rate. The feed was preheated by a heat exchanger using heavy raffinate product and again in the preheat section of the reboiler furnace, which used refinery fuel gas. Heavy raffinate was pumped from the bottom of the raffinate splitter tower and circulated through the reboiler furnace, where it was heated and then returned below the bottom tray. Heavy raffinate product was also taken off as a side stream at the discharge of the circulation pump and sent to storage. The flow of this side stream was controlled by a level control valve that, when placed in automatic mode, adjusted to maintain a constant level in the tower. The splitter tower was equipped with a level transmitter, which provided a reading of liquid level in the tower to the control room board operator[§§]. The transmitter measured the tower's liquid level in a 5-ft (1.5 m) span within the bottom 9 ft (2.7 m) of the 170-ft-tall (52 m) tower. The splitter tower also had two separate alarms that indicated high liquid level; one was programmed to sound when the transmitter reading reached 72% (a height of 7.6 ft or 2.3 m in the tower). The second, a redundant hardwired high-level switch was designed to sound when the liquid level reached 7.9 ft (2.4 m) in the tower

---

[§] A distillation tray has an opening at one end called a downcomer that allows liquid to flow down from the tray, and a series of bubble caps that allow the vapors to pass through from below. Hydrocarbon liquid enters the tray from the downcomer of the tray above. The liquid entering the tray is aerated with the vapor rising from the tray below that is flowing through the bubble caps to form froth on the tray. The froth flows across the tray unit it reaches as outlet weir. The froth flows over the weir into the downcomer, where the vapor is disengaged from the liquid.

[§§] This reading was also used by the computerized control board system to adjust the position of the level control valve.

(which corresponded to a reading of ~78% on the transmitter). The raffinate splitter was also fitted with a redundant low-level alarm.

The heavy raffinate product side stream flowed through two heat exchangers, one that exchanged heat from the heavy raffinate with the colder incoming feed to the raffinate splitter, and a second that cooled the heavy raffinate feed using water, before the heavy raffinate was sent to storage or blending tanks.

Light raffinate vapors flowed overhead and down a 148-ft-long (45m) section of pipe before they were condensed by the air-cooled fin fan condensers and then deposited into a reflux drum. Liquid from the reflux drum, called "reflux" was then pumped back up into the raffinate splitter tower above the tray (i.e., the top tray). The reflux drum was operated as a "flooded" drum, which means that during normal steady-state operation, it was kept completely full. The reflux drum also had high- and low-level alarms and safety relief valve set at 70 psig (483 kPa). A bypass line, which discharged into the raffinate splitter disposal header collection system, allowed release of non-condensable gas (e.g., nitrogen) and purge the system. During startup, uncondensed vapors that build-up in the drum were normally vented through a control valve to the refinery's 3-pound purge and vent gas system (Figure 19.198). This control valve malfunctioned and was not used during the March 23, 2005 startup.

Safety Relief Valves

To protect the raffinate splitter tower from overpressure, three parallel safety relief valves as shown in Figure 19.199 were located in the overhead vapor line 148 ft (45 m) below the top of the tower. The outlet of the relief valves was piped to a disposal header collection system that discharged into a blowdown drum fitted with a vent stack.

The set pressures on these relief valves were 40, 41, and 42 psig (276, 283, and 290 kPa), respectively (Table 19.64). An 8-in. NPS (8.625 in., 21.9 cm outer diameter) line, fitted with a manual chain valve, bypassed the safety relief valves and was used to release non-condensable gases and for system purging. The relief valves were designed to open and discharge primarily vapor into the raffinate splitter disposal header collection system when their set pressures were exceeded.

Disposal Header Collection Systems

The disposal header collection system received liquid and/or vapor hydrocarbons from venting relief and blowdown valves from equipment in the ISOM unit and discharged them to the blowdown drum. The header collection system included a 14-in. NPS (35.6 cm outer diameter) elevated pipe about 885 ft (270 m) long from the raffinate splitter tower (Figure 19.199). Other sections of the ISOM unit also discharged from two additional collection headers into the blowdown drum.

Blowdown Drum and Stack

The blowdown drum and stack were designed to accept mixed liquid and/or vapor hydrocarbons from venting relief and blowdown valves during unit upsets or following a unit shutdown. In normal operation, light hydrocarbon vapors disengage from the liquids, rise through a series of baffles, and disperse out of the top of the stack to the atmosphere. Any liquids or heavy hydrocarbon vapors released into the drum either fall, or condense and then fall, to the bottom of the drum, where they collect. Liquid would then be discharged from the base of the blowdown drum into the ISOM unit sewer system because a 6-in. NPS (6.625 in.; 15.24 cm outer diameter) manual block valve was chained open (Figures 19.200 and 19.201). This practice of discharging to the sewer was unsafe; industry safety guidelines recommend against discharging flammable liquids that evaporate readily into a sewer[§§§].

A liquid level, normally water, was maintained in the bottom of the blowdown drum. The height of this level was controlled by a "gooseneck" seal leg piped to a closed drain (Figures 19.200 and 19.201). A level sight glass was available to monitor the water level, and a high-level alarm was set to activate when the liquid level in the drum was close to flowing over the top of the gooseneck seal leg. A second manual block valve was located on a branch line of

---

[§§§] American Petroleum Institute (API) 1997. Guide for Pressure-Relieving and Depressuring Systems, API Recommended Practices 521, p52.

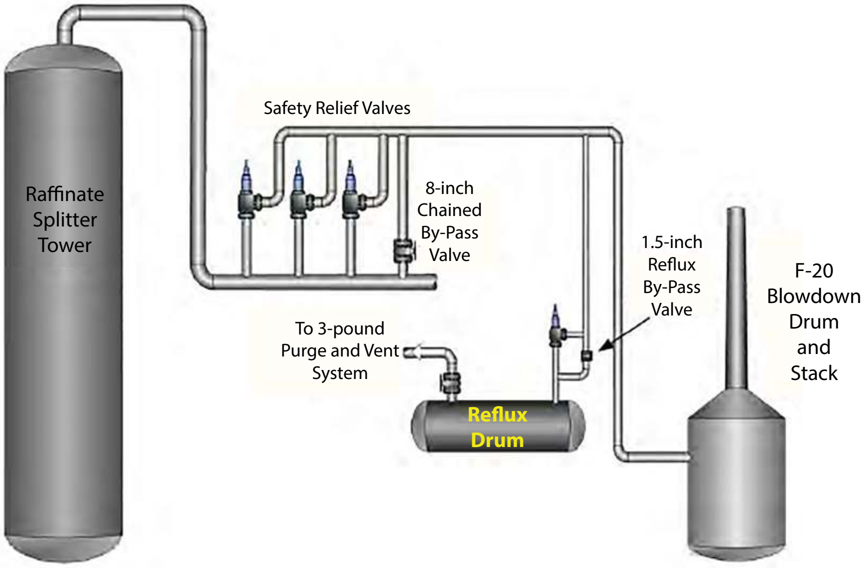


Figure 19.199 Disposal collection header system (Source: www.csb.gov).

Table 19.64 Specifications of safety relief valves installed on the raffinate splitter tower.

Safety relief valve	Original pressure setting (psig)	Present pressure setting (psig)	Inlet pipe diameter (in.)	Discharge pipe diameter (in.)	Valve ID
4P6	70	40	6	6	RV-1001GC
8T10	74	41	10	10	RV-1001GB
8T10	74	42	10	10	RV-1001GA

the blowdown drum discharge pipe (Figure 19.201). Following this valve, which was normally closed, was a manual steam-driven pump and a light slop tank*.

Incident Description

On the morning of March 23, 2005, during the startup of the raffinate splitter tower in the refinery's ISOM unit after a maintenance outage, operation's personnel pumped flammable liquid hydrocarbons into the tower for over three hours without any liquid being removed, which was contrary to startup procedure instructions. Critical alarms and control instrumentation provided false indications that failed to alert the operators of the high level in the tower. Consequently, unknown to the operations' crew, the 170-ft-tall (52 m) tower was overfilled, and liquid overflowed into the overhead pipe at the top of the tower.

The overhead pipe ran down the side of the tower to pressure relief valves located 148 ft (45 m) below. As the pipe was filled with liquid, the pressure at the bottom rose rapidly from about 21 to about 64 psi. The three pressure relief valves opened for six minutes, discharging a large quantity of flammable liquid to a blowdown drum with a vent stack open to the atmosphere. The blowdown drum and stack overfilled with flammable liquid, which led to a geyser-like release out the 113-ft-tall (34 m) stack. This blowdown system was an antiquated and unsafe design; it was originally installed in the 1950s and had never been connected to a flare system to safely contain liquids and combust flammable vapors released from the process.

---

* A slop tank is a tank designated for collecting equipment draining, tank washings, and other oily mixtures.

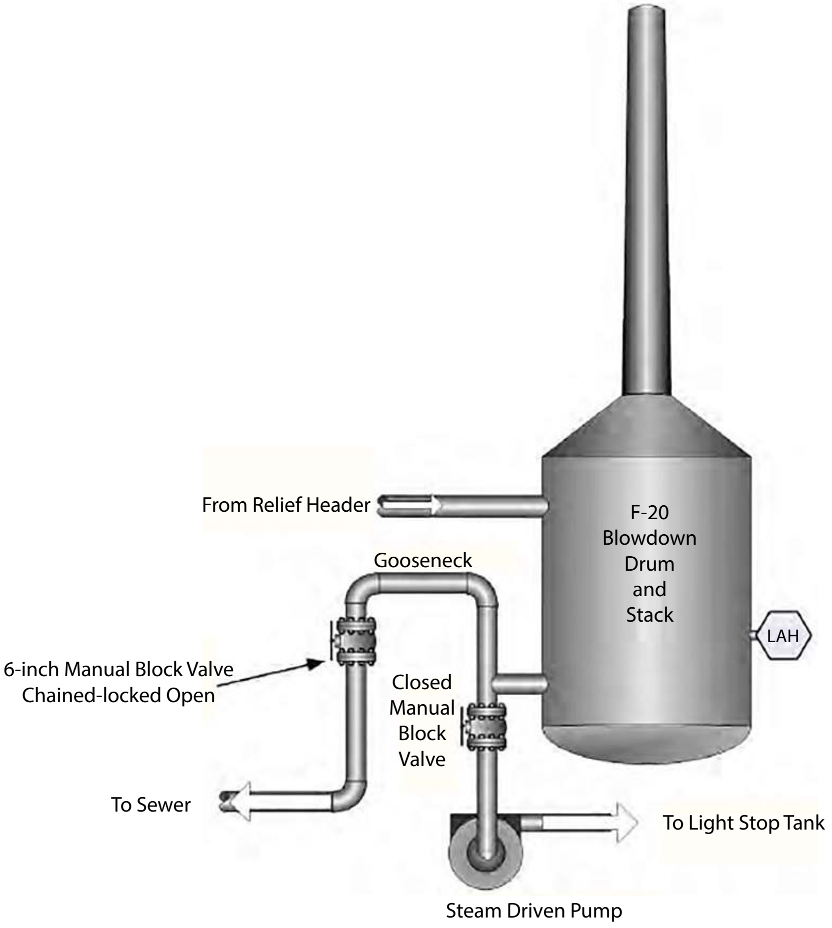


Figure 19.200 Blowdown drum and gooseneck (Source: www.csb.gov).

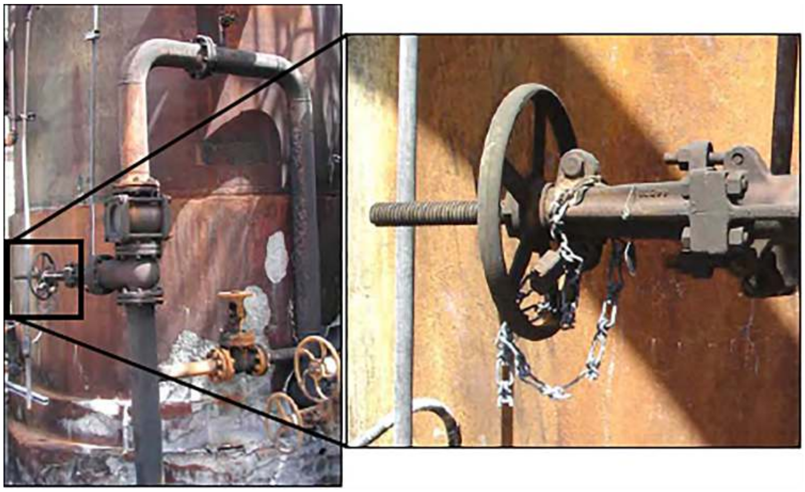


Figure 19.201 Left: Gooseneck piping and manual block valve (chained-locked open). Right: Close-up of manual block valve (Source: www.csb.gov).

Figure 19.202 Idling diesel pickup truck at north end of ISOM unit (Source: www.csb.gov).

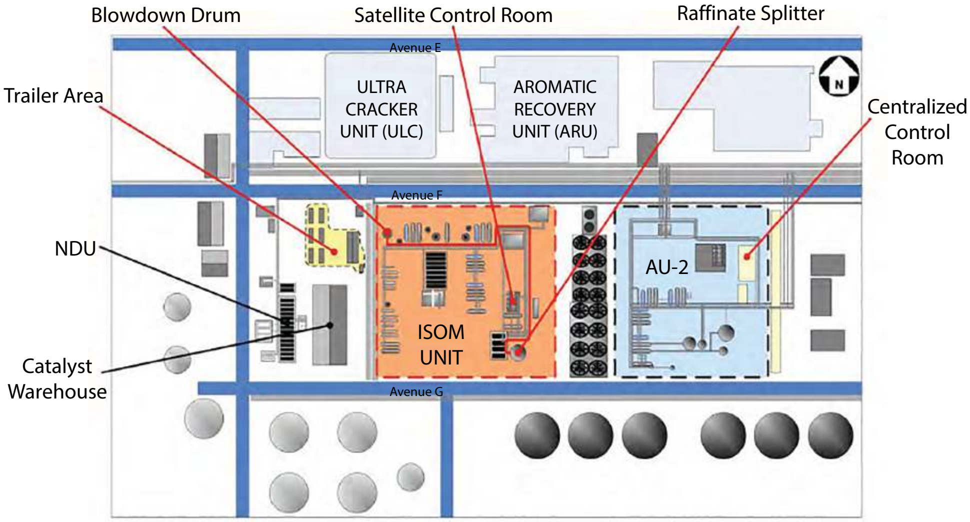


Figure 19.203 Refinery layout of the area surrounding the ISOM unit (Source: www.csb.gov).

The released volatile liquid evaporated as it fell to the ground and formed a flammable vapor cloud. The most likely source of ignition for the vapor cloud was a backfire from an idling diesel pickup truck located about 25 ft (7.6 m) from the blowdown drum (Figure 19.202). The 15 employees killed in the explosion were contractors working in and around temporary trailers that had been previously sited by BP as close as 121 ft (37 m) from the blowdown drum. Figure 19.203 shows the layout of the area surrounding the ISOM unit, and Figures 19.204 and 19.205 show the damage to the unit and the portable buildings, respectively.

Causes

The U.S. Chemical Safety and Hazard Investigation Board (CSB) investigation concluded "that while many departures to the startup procedure occurred, the key step that was instrumental in leading to the incident was the failure to establish heavy raffinate rundown to tankage, while continuing to feed and heat the tower. By the time the heavy raffinate flow was eventually started, the splitter bottoms temperature was so high, and the liquid level in the tower so high, that this intervention made matters worse by introducing significant additional heat to the feed (CCPS, 2008).

The investigation team concluded that the splitter was overfilled and overheated because "the shift board operator did not adequately understand the process or the potential consequences of his actions or inactions on March 23".

Figure 19.204 Portable buildings destroyed where contractors were located west of the blowdown drum (red arrow in upper left of the figure) (Source: www.csb.gov).

Figure 19.205 Texas City ISOM unit aftermath (Source: www.csb.gov).

The following underlying latent conditions contributed to the unsafe startup:

- A work environment that encouraged operations personnel to deviate from procedure.
- Lack of a BP policy or emphasis on effective communication for shift change and hazardous operations (such as unit startup).
- Malfunctioning instrumentation that did not alert operators to the actual conditions of the unit.
- A poorly designed computerized control system that hindered the ability of operations personnel to determine if the tower was overfilling.
- Ineffective supervisory oversight and technical assistance during unit startup.
- Insufficient staffing to handle board operator workload during the high-risk time of unit startup.
- Lack of a human fatigue prevention policy.
- Inadequate operator training for abnormal and startup conditions.
- Failure to establish effective safe operating limits.

Key Lessons

Many risk-based process safety elements were involved in the BP Texas City explosion. Five are listed below as taken from the US Chemical Safety Board (CSB) report.

Process Safety Culture

Process safety culture is the first of 20-risk based process safety elements (see Chapter 24, Volume 4). Perhaps, most striking of the CSB report are those with respect to the process safety culture at BP and the Texas City plant. Some of the CSB's findings regarding BP's process safety culture are listed below, as these findings could easily apply to other companies. The CSB recommended that BP created an "independent panel of experts to examine BP's corporate safety management systems, safety culture, and oversight of the North American refineries." This became known as the Baker Panel. The Baker Panel report focused on safety management systems at BP and resulted in 10 recommendations to the BP Board of Directors [220].

Selected CSB Findings

- "Cost-cutting failure to invest and production pressures from BP Group executive managers impaired process safety performance at Texas City.
- The BP Board of Directors did not provide effective oversight of BP's safety culture and major incident prevention programs. The Board did not have a member responsible for assessing and verifying the performance of BP's major incident hazard prevention program.
- Reliance on the low personal injury rate at Texas City as a safety indicator failed to provide a true picture of process safety performance and the health of the safety culture.
- A 'check the box' mentality was prevalent at Texas City, where personnel completed paperwork and checked off on safety policy and procedural requirements even when those requirements had not been met."

Selected Baker Panel Finding

- "BP has not instilled a common, unifying process safety culture among its U.S. refineries. Each refinery has its own separate and distinct process safety culture. While some refineries are far more effective than others in promoting process safety, significant process safety culture issues exist at all five U.S. refineries, not just Texas City. Although the five refineries do not share a unified process safety culture, each exhibits some similar weaknesses. The Panel found instances of a lack of operating disciplines, toleration of serious deviations from safe operating practices, and apparent complacency toward serious process safety risks at each refinery."

Process Knowledge Management

BP acquired the Texas City refinery as part of its merger with Amoco in 1999. Neither Amoco (the previous facility operator) nor BP replaced blowdown drums and atmospheric stacks, even though a series of incidents warned that this equipment was unsafe. In 1992, OSHA cited a similar blowdown drum and stack as unsafe, but the citation was withdrawn as part of a settlement agreement, and therefore, the drum was not connected to a flare as recommended. Amoco, and later BP, had safety standards requiring that blowdown stacks be replaced with equipment such as a flare when major modifications were made. In 1997, a major modification replaced the ISOM blowdown drum and stack with similar equipment, but Amoco did not connect it to a flare. In 2002, BP engineers proposed connecting the ISOM blowdown system to a flare, but a less-expensive option was chosen.

A poorly designed computerized control system display was found to be a cause in another major incident, the 1994 explosion and fire at a Texaco plant in Milford Haven, U.K. The control system did not calculate the material balance of the system, and the operators did not know how to make such calculations. The computerized control system was also configured to display only portions of the unit in discrete detailed sections and did not allow for a complete overview of the process, just like the control system screens for the Texas City ISOM unit. The U.K. Health and Safety

Executive (HSE) recommended that computerized control systems include a process overview and, as appropriate, material balance summaries to ensure full process oversight by operators (HSE, 1997).

Training and Performance Assurance

- A lack of supervisory oversight and technically trained personnel during the startup, an especially hazardous period, was an omission contrary to BP safety guidelines. An extra board operator was not assigned to assist, despite a staffing assessment that recommended an additional board operator for all ISOM startups.
- Supervisors and operators poorly communicated critical information regarding the startup during the shift turnover. BP did not have a shift turnover communication requirement for its operations staff. ISOM operators were likely fatigued from working 12-hour shifts for 29 or more consecutive days.
- The operator training program was inadequate. The central training department staff had been reduced from 28 to 8, and simulators were unavailable for operators to practice handling abnormal situations, including infrequent and high hazard operations such as startups and unit upsets.

Management of Change (MOC)

Changes made to processes, equipment, procedures, buildings, and personnel at the refinery were not systematically reviewed to ensure that an adequate margin of safety was maintained.

The refinery required that an MOC be initiated for any change that involved process chemicals, process/equipment technology, equipment piping, process control and instrumentation, operating procedures, safe operating limits, relief/safety systems, personnel/staffing/organizations/outsourcing, and occupied buildings. The program revolved around the refinery asset superintendents, who were responsible for most aspects of the program. Support for the MOC program was provided by the refinery's PSM group, which managed the electronic MOC database, rained PHA leaders, and tracked resolution of MOC action items.

- BP Texas City did not effectively access changes involving people, policies, or the organization that could impact process safety. For example, the control room staff was reduced from 2 people to one, who was overseeing three units.
- Local site MOC rules, required that where a portable building was to be placed within 350 ft (100 m) of a process unit a facility sitting analysis had to be carried out. However, this location had already been used many times for these trailers. Not performing an effective MOC put all the people in the portable buildings at unnecessary risk.

Asset Integrity and Reliability

- The process unit was started despite previously reported malfunctions of the tower level indicator, level sight glass, and a pressure control valve.
- Deficiencies in BP's mechanical integrity program resulted in the "run to failure" of process equipment at Texas City.

Pressure Relief System Design Basis and Disposal System Design

Texas City management did not identify all credible overpressure scenarios, fully document potential overpressure events, or calculate relief flow rates for all the potential overpressure scenarios on the raffinate splitter. Without this information, Texas City management could not ensure that the blowdown disposal system was safely designed. The lack of design basis documentation also points to a failure of PSM systems, such as process safety information and PHA requirements.

The CSB reviewed the BP design basis calculations for the installed raffinate splitter tower relief system, and had a contractor perform emergency relief design calculations on the raffinate splitter tower to determine if it met the requirements of the American Society of Mechanical Engineers (ASME) "Boiler and Pressure Vessel Code, Section VIII – Unfired Pressure Vessels," and other applicable industry guidelines, such as API 521 (see Chapter 23, Volume 4 of this series).

The computer program incorporated the AIChE; DIERS, technology for both viscous and non-viscous flashing, frozen, and hybrid two-phase flows through emergency relief devices and pipes. Also included were routines to calculate inlet pipe pressure drop and discharge pipe backpressure for choked and subcritical vapor, two-phase, and sub-cooled liquid flows from safety valves and breather vents. The program reported choked or unchoked and turbulent or laminar flow through each device and associated inlet or discharge pipe, and the first occurrence of safety valve inlet pipe high pressure drop or discharge pipe high backpressure. Warning messages were provided, if the emergency relief devices cannot maintain the system pressure to meet ASME Pressure Vessel Code and API requirements (ASME, 2001 [221]; API, 2000 [222], 1997 [223], and 1994 [224]).

Based on a review of records from BP, the original design basis of the safety relief valves installed on the raffinate splitter tower was the heat input at the production capacity and flow at an accumulation of 1.16 (process upset/multiple relief devices) of the 76 psig (483 kPa) Maximum Allowable Working Pressure (MAWP). When the MAWP of the raffinate splitter tower was derated to 40 psig (276 kPa), the area of the safety relief valves was not increased to reflect the flow at the lower set pressure, nor was the production capacity of the unit reduced. Neither does documentation exists showing that the maximum flow rate of the fuel available to the raffinate splitter feed preheater/fractionator reboiler was reduced.

The relief system design and design basis (controlling case) for the raffinate splitter tower with a MAWP of 40 psig (276 kPa) is a reflux failure, with a worst credible scenario total loss of condenser cooling at the design capacity. Unfortunately, BP used a partial loss of cooling with reflux, rather than the total reboiler duty with no reflux for the recent calculations, resulting in a design error [220]. The original emergency relief system design for the raffinate splitter with a 70 psig (483 kPa) MAWP does not appear to contain this error.

The raffinate splitter reboiler duty is rated as 59.6 MM Btu/h; the raffinate splitter overhead condenser duty is rated as 59.4 MM Btu/h. Using the latent heat vaporization range of 113.3 Btu/lb, the required capacity of the installed emergency relief at an accumulation of 1.16 for the 40 psig (276 kPa) MAWP would be 525,215 pph. Table 19.64 shows three consolidated bellows-type safety relief valves to provide the required emergency relief for the raffinate splitter system with the current MAWP of 40 psig (276 kPa), which was derated from 70 psig (483 kPa) in 2003 due to corrosion considerations.

The capacity and backpressure on these safety relief valves were evaluated using a 0.95 coefficient of discharge appropriate for a consolidated bellows safety relief valve, an ASME code capacity reduction factor of 1.0, and the API 520 consensus K_b factor to account for the decreasing lift and resultant flow reduction of bellows safety relief valves as the discharge pipe backpressure increases. The flow capacities were evaluated at 1.16 MAWP values with set pressures of 40 and 70 psig (276 and 483 kPa) (Table 19.65).

Table 19.65 shows that excessive backpressure of 60.8% of the safety relief valve set pressures were found for vapor flow from the bellows safety relief valves for a MAWP of 40 psig (276 kPa) and 61.8% for a MAWP of 70 psig (483 kPa). These high backpressures show that the diameter of the bellows safety relief valves discharge pipe header should have been 20 in. rather than 14 in., for either the 40 or the 70 psig (276 or the 483 kPa) MAWP designed calculations.

The ASME Code, Appendix M, Para. M-8 states "The sizing of any common discharge header downstream from each of two or more pressure relieving devices that can reasonably be expected to discharge simultaneously shall be based on the total of their outlet areas…" This design error should have been caught by inspection, during the many PHA revalidations, as per of the OSHA PSM "relief system design and design basis" calculations or when the pressure relief calculations were completed that supported lowering the MAWP of the raffinate splitter.

Table 19.65 Flow from the safety relief valves on the raffinate splitter.

No discharge pipe header		**14-in. discharge pipe header**		**20-in. discharge pipe header**	
Tower (top-mounted SRVs)	**Flow rate, pph**	**Flow rate, pph**	**% Backpressure**	**Flow rate, pph**	**% Backpressure**
Vapor flow (40 psig MAWP)	416,467	213,339	60.8	378,862	37.0
Vapor flow (70 psig MAWP)	637,441	316,194	61.8	542,942	38.0

BP has not provided relief system design and design basis backpressure calculations for the safety relief valve discharge pipe header from the raffinate splitter to the blowdown drum, as required by the OSHA PSM standard [29 CFR 1910.119(d)(3)(i)(D)].

The sizing information for the safety relief valve protecting the raffinate splitter reflux drum reflects only vapor flow. The reflux drum operation changed to operate liquid-full. The safety relief valve should have been sized for a subcooled liquid or flashing two-phase vapor-liquid flow instead of vapor [220].

Lieberman [226] provided a plausible scenario as to why the operators misread the level gauge in the raffinate splitter and highlighted that the fundamental causes of the explosion are as follows:

- Relief valve is not connected to plant flare.
- Operator laxity.

1. The design of the loop seal draining the blowdown tower to the 3-psig condensate collection system.
2. Elevation of the splitter relief valves relative to the top of the tower.
3. Lack of technical support during startup operations.
4. Operators not understanding the relationship between tower bottom temperature and tower bottoms indicated level.
5. Operators not understanding the concept of a level-troll being "tapped-out".
6. Failure to check liquid head pressure at the bottom of the tower.

Readers can visit the US Chemical Safety Board website to view the animated video of BP American refinery explosion in Texas City (www.csb.gov/bp-america-refinery-explosion).

Case Study 2

Chevron Richmond Refinery # 4 Crude Unit, Richmond California, USA, August 6, 2012 [219]

On August 6, 2012, The Chevron U.S.A., Inc. Refinery in Richmond, California experienced a catastrophic pipe rupture in number 4 crude unit. The incident occurred from the piping referred to as the "4-sidecut" stream, one of the several process streams exiting the refinery's C-1100 crude unit atmospheric column. The pipe ruptured occurred on a 52-in.-long component of the 4-sidecut 8-in. line (the 52-in. component*). At the time of the incident, light gas oil was flowing through the 8-in. line at a rate of ~10,800 barrels per day (bpd).

The ruptured pipe released flammable, high-temperature light gas oil, which partially vaporized into a large, opaque vapor cloud that engulfed 19 Chevron U.S.A. Inc. employees. At 6:33 p.m., approximately two minutes following the release, the released process fluid ignited. Eighteen of the employees safely escaped from the vapor cloud just before ignition; one employee, a Chevron refinery firefighter, was inside a fire engine that was caught within the fireball when the process fluid ignited. He escaped due to his protective equipment, but six Chevron employees suffered minor injuries during the incident and subsequent emergency response efforts.

---

* The term "component" refers to a portion of piping between welds or flanges. It includes straight run piping and pipe fittings.

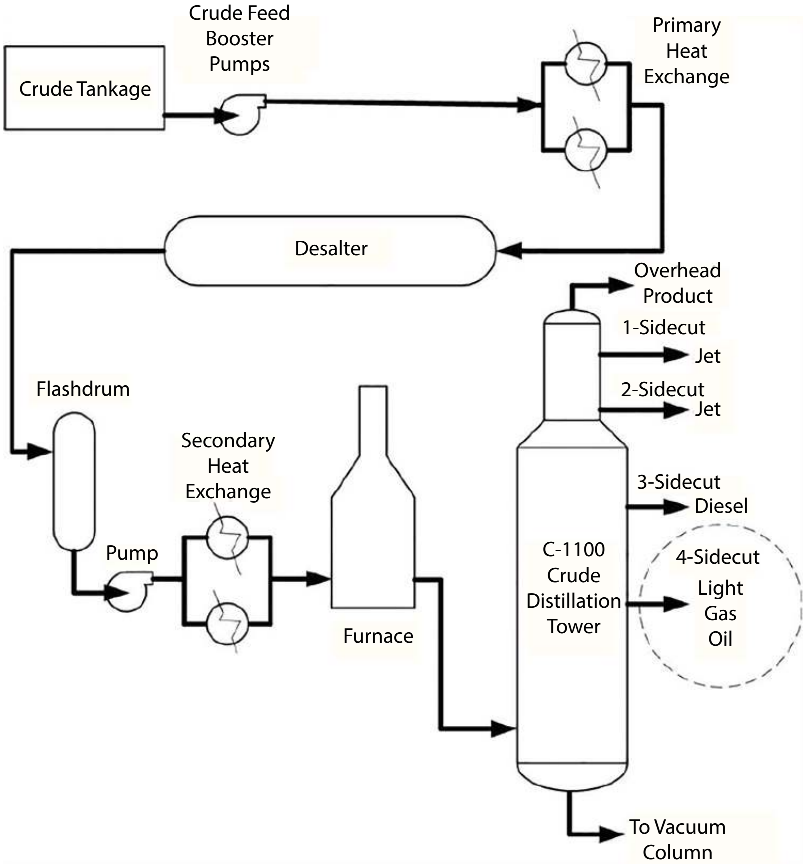


Figure 19.206 Schematic of C-1100 crude unit atmospheric column and upstream process equipment (Source: www.csb.gov).

The release, ignition, and subsequent burning of the hydrocarbon process fluid resulted in a large plume of vapor, particulates, and black smoke, which traveled across the surrounding area. The incident was finally brought under control at 11:12 p.m. In weeks following the incident, ~15,000 people from the surrounding communities sought medical treatment at nearby medical facilities for ailments, including breathing problems, chest pain, shortness of breath, sore throat, and headaches. Approximately, 20 of these people were admitted to local hospitals as inpatients for treatment.

Description of the Process

The Richmond California Chevron Refinery's # 4 crude unit performs the initial processing step in the refining process. Raw crude oil stored in storage tanks is pumped to the crude unit. After an initial "cleaning" of the oil using a desalter, which removes corrosive salts, solids, and water**, the oil is preheated and enters the C-1100 Crude unit atmospheric column at approximately 675°F (357°C). The crude column separates through distillation various hydrocarbons component mixtures in the crude feed into multiple streams coming from the column with differing boiling points. These streams include an overhead light hydrocarbon stream, jet oil streams, a diesel stream, a light gas oil stream, and a bottoms stream composed of heavy liquid hydrocarbons. Each stream is further refined and processed in subsequent units within the refinery.

The Incident

On August 6, 2012, the incident occurred from the piping referred to as the "4-sidecut" line, one of the several process streams exiting the crude column (Figure 19.206)***. Figure 19.206 shows the light gas oil, the crude unit

---

** Removing chloride salts and water prevents the formation of hydrochloric acid, which can severely corrode downstream equipment. Other salts and solids are removed to prevent fouling within equipment such as heat exchangers, which can significantly reduce heat transfer.

*** The atmospheric column separates crude oil feed into different streams through distillation. These streams are further processed in other units in the refinery.

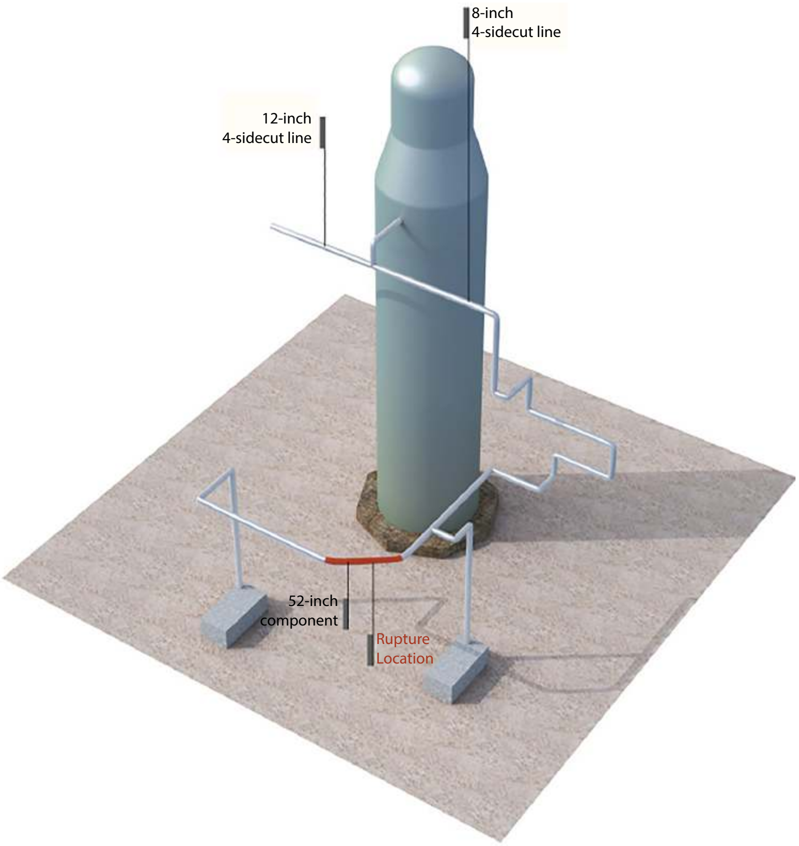


Figure 19.207 4-sidecut line configuration and rupture location (Source: www.csb.gov).

Figure 19.208 CSB amination depicting operator identifying the leaking 4-sidecut pipe (Source: www.csb.gov).

Figure 19.209 Photo of rupture on 4-sidecut 52-in. component (Source: www.csb.gov).

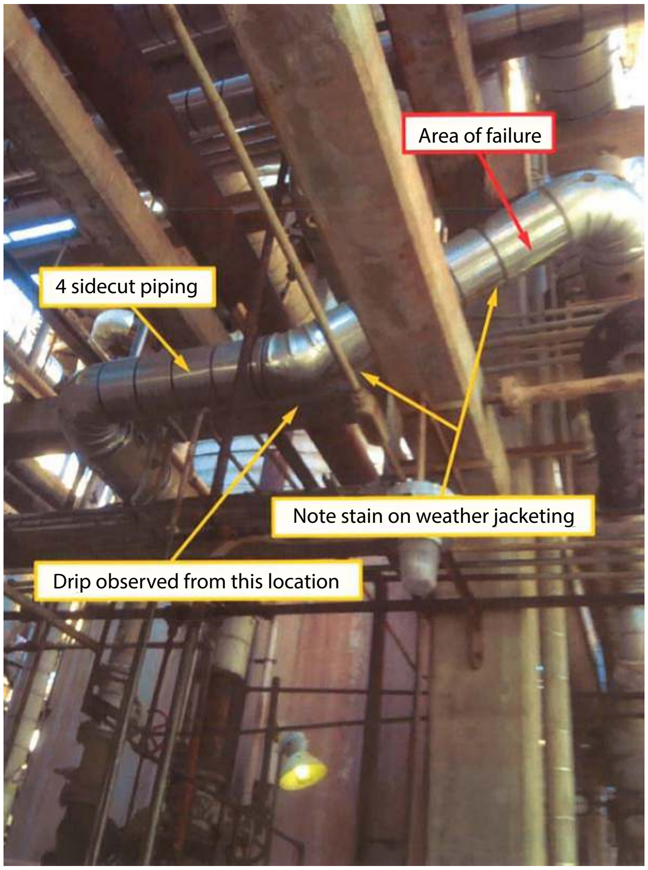


Figure 19.210 Photo taken of the leaking 4-sidecut pipe on August 6, 2012, at the Chevron Richmond Refinery. Insulation obscured the actual leak location. Stain signifies where hydrocarbon process fluid was leaking from the 4-sidecut piping (Source: www.csb.gov).

4-sidecut process fluid, exits the atmospheric column via a 20-in. nozzle, and is split into a 12-in. line and an 8-in. line. The pipe rupture (Figure 19.207) occurred on a 52-in.-long component of the 4-sidecut 8-in. line (the 52-in. component). The line operated at a temperature near 640°F[†] and had an operating pressure of approximately 55 psig at the rupture location. At the time of the incident, light gas oil was flowing through the 8-in. line at a rate of approximately 10,800 bpd[§].

At approximately 3:50 p.m., on August 6, 2012, an outside operator performing routine checks of piping and equipment found an 18-in. puddle of what appeared to be a diesel-like material on the refinery concrete pad (Figure 19.208). Identifying that the leak was occurring from overhead, the operator observed intermittent drips as they accumulated on the underside of an insulated pipe 14 ft above ground level. The leaking pipe was identified to be a portion of the 4-sidecut piping that originated on the crude column. Visually analyzing the piping, the operator determined that the line could not be isolated from the process. The operator's supervisor arrived at the leak location, shortly followed by the shift team leader. These personnel observed that the leak was dripping at a rate of approximately 40 drips per minute. The piping was insulated, so the individuals gathered near the leak could not identify its precise source (Figure 19.209). They concluded that the leak was not significant enough to require a shutdown but was still a serious situation. Shortly after 4:00 p.m., they called the Chevron Fire Department to the scene, a typical practice at the refinery when leaks are discovered. Firefighters began to arrive at approximately 4:07 p.m. and established an incident command structure. A hot zone of 20 ft by 20 ft was established and taped off around the leak location by the incident commander. The area outside of the hot zone was considered the cold zone or safe zone[§§].

Beginning at approximately 4:15 p.m., many additional personnel were called to the scene of the leak to assist in the leak analysis. Various operation's personnel were called to the leak. Two Chevron inspectors reported to the leak location to provide information on inspection history of the 4-sidecut line. The lead crude unit process engineer also arrived at the leak location to determine an estimate on the hole size and the quantity of material leaking so that proper environmental release calculations could be performed.

At approximately 5:00 p.m., the shift team leader left the scene of the leak and went to the control room. He directed the board operator to reduce the feed to the 4-sidecut line by 5,000 bpd. Ultimately, a large group of Chevron employees—40 people—accumulated at the leak location. They began discussing their options to mitigate or stop the leak. The inspectors informed the group that the 4-sidecut pipe walls (Figure 19.210) were thinning due to sulfidation corrosion (see Chapter 24 of Volume 4 of this series), but data collected as recently as two months prior indicated the 4-sidecut line had sufficient wall thickness to last until the next turnaround in 2016. This assessment led the group to believe that a localized mechanism, such as abrasion on the line from a pipe support near the dripping location, was the likely cause of the leak. The group then called the leak repair contractor to the leak location to assess the possibility of clamping the line to stop the leak. Figure 19.211 shows a photo of a typical leak repair clamp.

The group then decided to remove the insulation from the 4-sidecut pipe to determine the cause of the leak. The first attempt to remove the insulation was made by pulling on the insulation bands from the ground using a pike pole, which was unsuccessful. Contractors arrived at the scene to build the scaffold as shown in Figure 19.212. Once the scaffolding was built, two firefighters were directed to climb the scaffold and remove the aluminum sheathing and insulation from the 4-sidecut pipe. As the firefighters were removing the sheathing of the 4-sidecut line (Figure 19.213), white hydrocarbon vapor visibly began to emerge from under the now-exposed insulation material. The firefighters continued to remove the sheathing despite the formation of the hydrocarbon vapor. During the

---

[†] The autoignition temperature for this process, the temperature at which a material will combust in the presence of sufficient oxygen without an ignition source, was 640°F. This number is based on the Chevron Light Gas Oil Material Safety Data Sheet. Chemical testing of 4-sidecut samples following the incident indicated lower autoignition temperatures; however, these samples may not have been representative of typical 4-sidecut process fluid. Chevron instrumentation indicates that the process fluid entered the 4-sidecut piping at a temperature near 640°F and cooled at 625°F before reaching the piping circuit pumps downstream of the rupture location.

[§] This rate is the equivalent of 315 gpm. A barrel equals 42 gal.

[§§] A decontamination corridor is often established in the warm zone, an area established between the hot zone and the cold zone. "Decontamination involves thorough washing to remove contaminants. It should be performed in an area upwind of the hot zone. An area that is uphill, with good drainage, and easily accessible for responders is preferred."

Figure 19.211 Example of a leak repair clamp for piping. It is installed over the leak location to prevent process fluid leakage to the atmosphere (Source: www.csb.gov).

Figure 19.212 CSB amination of contractors erecting scaffolding beneath the leak location (Source: www.csb.gov).

continued sheathing removal, insulation that was soaked with hot 4-sidecut hydrocarbon autoignited once exposed to oxygen-only feet from the firefighters. The hose teams immediately put out the fire, and both firefighters quickly came down from the scaffold (Figure 19.214). The firefighters on the scaffolding successfully removed much of the aluminum sheathing surrounding the insulation; however, underlying insulation still obscured the location of the leak. Using the fire hoses to knock the insulation off the pipe, the team knocked off the insulation up to the location where the aluminum sheathing had been removed. At this point, they realized that the leak had significantly worsened; hydrocarbon liquid was now spraying from the pipe. The operational managers at the site decided to shut the unit down, an action that requires hours to complete.

Consequences

A vapor cloud quickly began to accumulate. The hose teams attempted to keep the cloud at bay by spraying it with firefighting water. Suddenly, the vapor cloud worsened, engulfing 19 firefighters and operators standing in both the hot zone and cold zone in the hot hydrocarbon cloud. The cloud was dense and very hot, and many of the individuals caught in the cloud were not able to see anything around them. Each person engulfed in the cloud began working

Figure 19.213 CSB amination of firefighters attempting to remove the 4-sidecut insulation, the resulting fire and fire extinguishing (Source: www.csb.gov).

Figure 19.214 CSB animation of firefighters who dropped to their hands and knees to escape the vapor cloud (Source: www.csb.gov).

Figure 19.215 CSB animation of firefighter who was inside the fire engine when the light gas oil ignited (Source: www.csb.gov).

their way out of the vapor cloud, and the firefighters operating the two hoses dropped to their knees to follow their hose lines to safety, feeling their way out of the cloud (Figure 19.214).

At approximately 6:30 p.m., two minutes after the large vapor cloud formed, the light gas oil ignited. Eighteen employees safely escaped from the cloud just before ignition. One employee, a firefighter, was inside a fire engine that was engulfed in the fireball when the light gas oil ignited (Figures 19.215 and 19.216). He was able to escape through the flames surrounding the fire truck and made his way to safety without physical injury because of his heavy full body firefighting protective equipment that he wore.

The leak resulted in a large plume of vapor which traveled across the surrounding area. The ignition and subsequent burning of the hydrocarbon process fluid created a large black cloud of smoke, which also swept across the surrounding area (Figure 19.216). The fire was brought under control at 11:12 p.m., and in the weeks following the incident, nearby medical facilities received over 15,000 members of the public seeking treatment for ailments, including breathing problems, chest pain, shortness of breath, sore throat, and headaches.

Key Findings

The CBS commissioned Anamet, Inc., a material engineering and laboratory testing company to conduct testing of the 4-sidecut pipe, including the failed 52-in. component. The testing concluded that the rupture was due to pipe wall thinning caused by sulfidation corrosion (See Chapter 24 in Volume 4 of this series).


Photo credit: Fototaker.net
Photo credit: Fototaker.net
Photo credit: Fototaker.net
Photo credit: Fototaker.net

Figure 19.216 Initial vapor cloud formation (white cloud) and subsequent ignition (black smoke) as seen from a pier in San Francisco, California (Source: www.csb.gov).

Sulfidation corrosion, also known as sulfidic corrosion, is a damage mechanism that causes thinning in iron-containing materials, such as steel, due to the reaction between sulfur compounds and iron at temperatures ranging from 450°F to 1,000°F (232°C–538°C). For pipe walls, this damage mechanism causes gradual thinning over time. Sulfidation corrosion is common in crude oil distillation, where naturally occurring sulfur and sulfur compounds found in crude oil feed, such as hydrogen sulfide (H_2S) are available to react with steel piping and equipment. Process variables that affect corrosion rates include the total sulfur content of the oil, the sulfur species present, flow conditions, and the temperature of the system. All crude oil feeds contain sulfur compounds; therefore, sulfidation corrosion is a damage mechanism present at every refinery that processes crude oil. Sulfidation corrosion can cause thinning to the point of pipe failure when not properly monitored and controlled.

Sulfidation corrodes carbon steel at a much faster rate in comparison with its effect on other materials of construction, such as steels with a higher chromium content. In addition to its naturally faster rate of sulfidation corrosion when compared with higher chromium steels, carbon steel can also experience significance variation in corrosion rates due to variances in silicon content, a component used for the steel manufacturing process. Carbon steel piping containing silicon content less than 0.10% weight can corrode at accelerated rates up to 16 times faster than carbon steel piping containing higher percentages of silicon. Figure 19.217 shows how carbon steel corrosion rates can greatly vary depending on silicon content.

The Chevron Richmond Refinery 4-sidecut piping circuit containing the 52-in. component that failed was constructed of ASTM A53B carbon steel, which had no minimum specification for silicon content. Post-incident testing of samples of the 4-sidecut piping from the Chevron Richmond Refinery identified silicon content ranging from 0.01 to 0.2 wt %. Twelve samples were taken from the 8-in. and the adjacent 12-in. 4-sidecut line; six had a silicon concentration of less than 0.10 wt % (Figures 19.218 and 19.219). The 52-in. pipe component that ruptured on the day of the incident had a silicon content of only 0.01 wt %. This shows the inherent variability in ASTM A53B carbon steel sulfidation corrosion rates, the elbow component directly upstream of the 52-in. component that failed had a silicon concentration of 0.16 wt % and showed considerably less thinning (Figure 19.220).

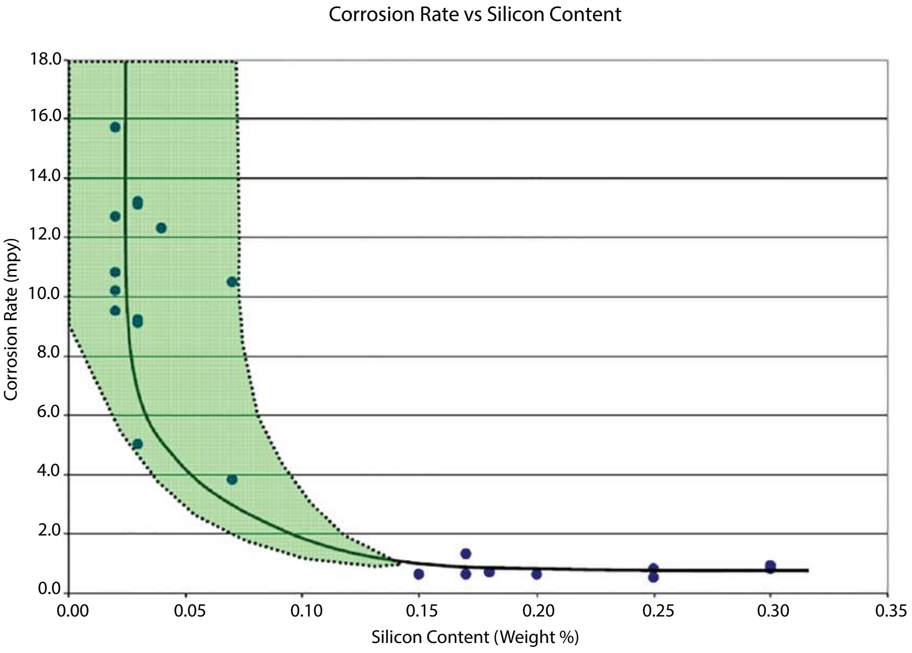


Figure 19.217 Graph of sulfidation corrosion rates with respect to silicon content in carbon steel. This graph shows how corrosion rates increase in carbon steel containing decreasing percentages of silicon (check: API RP 939-C.) (Source: www.csb.gov).

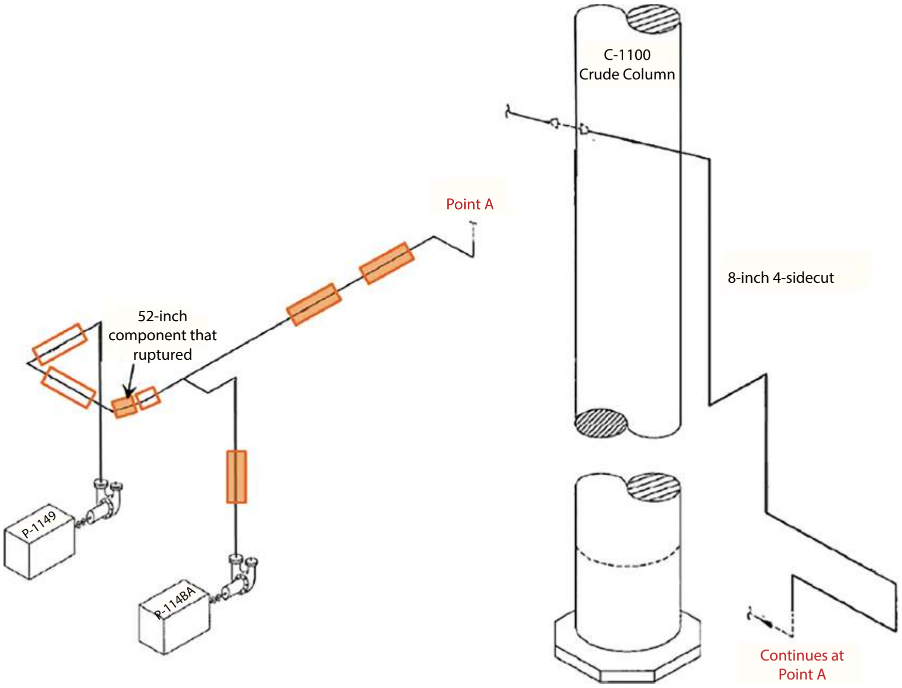


Figure 19.218 Location of metallurgical samples taken from 8-in. 4-sidecut piping post-incident. The seven samples taken are boxed in orange. Four of the seven components sampled (shown with an orange fill) were found to have a silicon content less than 0.10 wt % (Source: www.csb.gov).

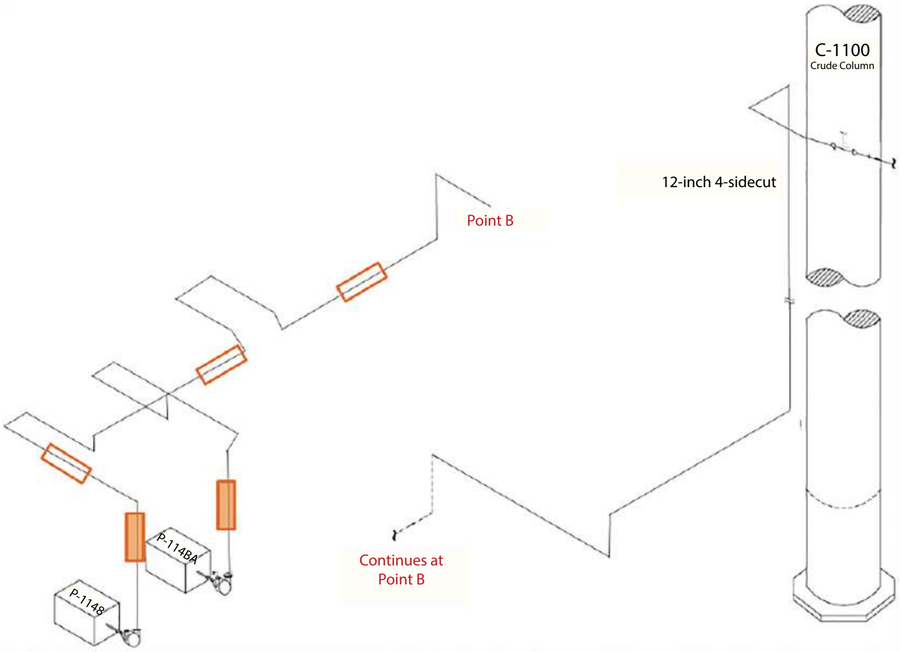


Figure 19.219 Locations of metallurgical samples taken from 12-in. 4-sidecut piping post-incident. The five samples taken are boxed in orange. Four of the five components sampled (shown with an orange fill) were found to have a silicon content less than 0.10 wt % (Source: www.csb.gov).

Recommendations

The CSB recommended that Chevron management, engineers, inspectors, and operators see the importance of having good process safety systems and the value of ensuring that work processes are safe, and equipment is reliable. A desire to be safe is inadequate; to ensure process safety, organizations must have been well-designed, integrated system, rigorous programs, and strong leadership for these programs.

Chevron must ensure the effectiveness of implementing new safety-critical programs at the refinery level, such as the Chevron Energy Technology Company (ETC) Sulfidation Failure Prevention Initiative, by developing a formalized system that identifies one individual or group with decision-making authority within each refinery to be responsible and accountable for program implementation.

Chevron should implement a framework for tracking leading and lagging indicators to continuously monitor and improve process safety, as an indicator could have ensured that the ETC Sulfidation Failure Prevention Initiative and its status at the Chevron Richmond Refinery were at the forefront of management's attention. This includes an indicator into the Operational Excellence and Reliability Intelligence (OERI) system for tracking the implementation of a key ETC process safety recommendations or new industry guidance will aid in preventing future incidents at Chevron refineries.

The CSB recommended that California enhances and restructures its process safety management (PSM) regulations for petroleum refineries by including specific goal-setting attributes. The recommendations included language to strengthen stop work authority, "The regulation should provide workers and their representatives with the authority to stop work that is perceived to be unsafe until the employer resolves the matter, or the regulator intervenes."

Facilities should develop an emergency response plant that "shall be developed and implemented to handle anticipated emergencies prior to the commencement of emergency response operations." Pursuant to this requirement, Chevron developed an emergency response plan comprised of several refinery instructions outlining the required response activities for specific types of emergencies. For example, the Chevron Richmond Refinery developed refinery instructions to be followed in the case of a fire in the refinery, oil spills to the adjacent bay and release of hydrogen

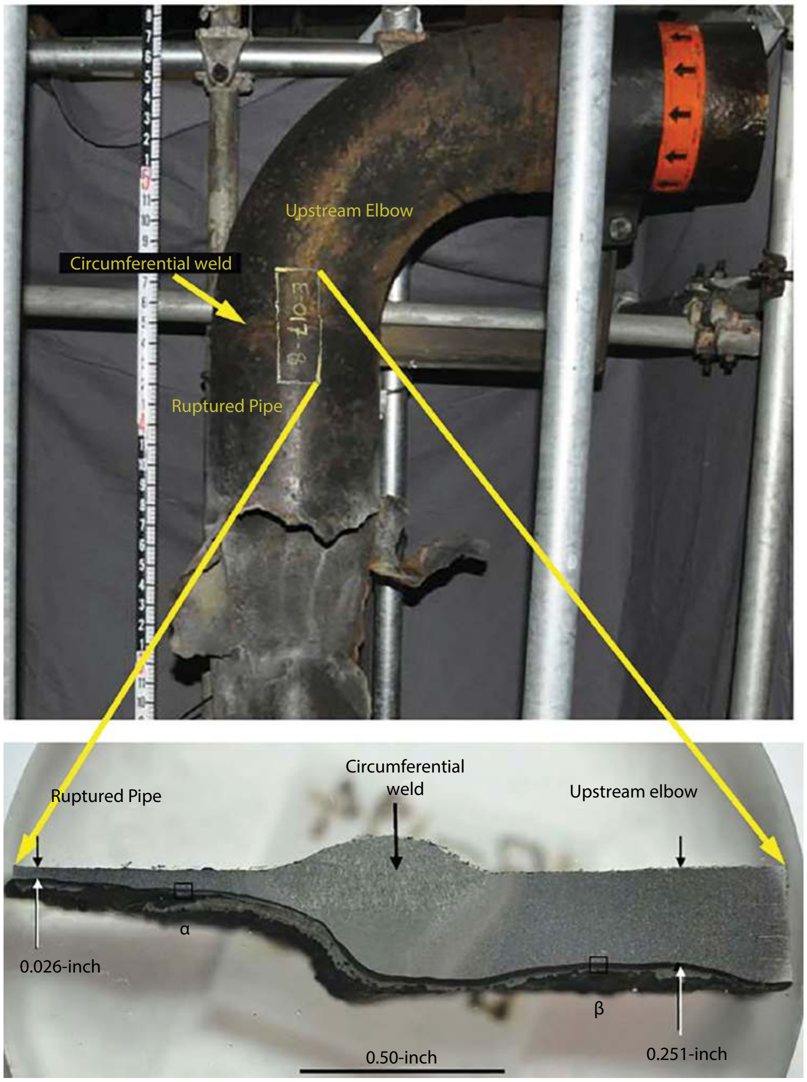


Figure 19.220 4-sidecut piping sample (E-017-8) analyzed by Anamet Labs showing the relative thickness of low silicon piping on the left and the high silicon piping on the right. The ruptured 52-in. pipe component (left) contained 0.01 wt % silicon, and the upstream elbow component (right) contained 0.16 wt % silicon. The initial nominal thickness of this piping was 0.322 in. (Source: www.csb.gov).

sulfide, among other emergency situations. The CSB analysis suggests that had an effective leak response protocol been in place during the August 6, 2012, incident, it likely would have been clear there was little to no chance that the 4-sidecut leak could be stopped and that the crude unit should not continue to operate for any significant period. The analysis of the end result for every potential leak mitigation action likely would have resulted in the same decision: to shut down the crude unit immediately.

If Chevron had used a pre-established response procedure that required consultation with various subject matter experts (Figure 19.221), the Incident Commander[§§§] could have identified that sulfidation-induced failures have historically resulted in large blowouts and catastrophic failures. A clamp to mitigate a sulfidation leak (Figure 19.211) would thus have to encompass the entire affected piping segment.

[§§§] Chevron defines the Incident Commander as the "senior emergency response official" at an incident site who is responsible for overall incident objectives and controlling emergency operations at the site. Additional responsibilities include site hazard assessment to the extent possible all hazardous substances or conditions present are identified, establish hot zone, and address exposure control and PPE selection."

Figure 19.221 Ideal communication flow to Incident Commander during refinery process fluid leak incident. In industrial process fluid leak emergency situations, it is essential that various personnel with different areas of expertise communicate their relevant knowledge to the Incident Commander (Source: www.csb.gov).

Industry Sulfidation Corrosion Guidance

Industry organization and trade associations, such as the American Society of Mechanical Engineers (ASME), the National Fire Protection Association (NFPA), and the American Petroleum Institute[ξ] (API), develop codes, standards and recommended practices which define requirements and recommendations to conduct operations safety. Codes, standards, and recommended practices are developed by a committee of experts based on consensus and are often updated on fixed-year intervals. Codes can be adopted as requirements by regulatory agencies or authorities having jurisdiction. In addition, since these requirements are often considered recognized and generally accepted good engineering practices (RAGAGEP), regulators can cite industrial facilities for not following them.

Codes and standards developed by API give specific information and guidance to industry on the technical details of sulfidation corrosion and ways to inspect piping and equipment susceptible to it. However, the CSB identified significant gaps in these standards. There is varying, sometimes conflicting information in many of API's standards and recommended practices that describe sulfidation corrosion. All these publications should align to deliver a constant message to users on inspecting for sulfidation corrosion and preventing sulfidation failures in low-silicon carbon steel.

API RP 939-C: Guidelines for Avoiding Sulfidation (Sulfidic) Corrosion Failures in Oil Refineries

API RP 939-C: Guidelines for Avoiding Sulfidation (Sulfidic) Corrosion Failures in Oil Refineries is the primary industry guidance document on ways to monitor and control sulfidation corrosion. It aims "to provide practical guidance to inspectors, maintenance, reliability, project, operations, and corrosion personnel on how to address sulfidation corrosion in petroleum refining operations[ξξ]". It was published in 2009 following a string of sulfidation corrosion-related incidents in the early 2000s.

[ξ] American Petroleum Institute: Procedures for Standards Development. 4th ed., 2009, Section 5.4. See http://www.api.org/publications-standards-and-statistics/~/media/Files/Publications/FAQ/2011-Procedures-Final.ashx (accessed September 15, 2014).

[ξξ] API RP 939-C: Guidelines for Avoiding Sulfidation (Sulfidic) Corrosion Failures in Oil Refineries, Introduction, 2009.

The recommended practice cautions that low-silicon carbon steel piping can corrode at an accelerated rate. It states that carbon steel will appear to be of sufficient thickness based upon measured corrosion rates, typically at condition monitoring locations (CMLs) placed on elbows and fittings with higher silicon content, until failure occurs at an unmonitored or unidentified low-silicon piping component.

API RP 939-C specifically discusses risks associated with sulfidation corrosion in low-silicon carbon piping. It acknowledges that older carbon steel piping can have low silicon content, creating a major inspection challenge, because small piping sections (pups) or fittings with low silicon may corrode at rates 2 to 10 times faster than surrounding higher silicon piping. Unless the refinery is fortunate enough to have located an inspection point on that particular section of pipe or fitting, it is very difficult to detect the thinning component.

The document also communicates the risk of sulfidation corrosion failures, stating "ruptures are possible leading to the potential release of large quantities of hydrocarbon streams, and sulfidation corrosion continues to be a significant cause of leaks leading to equipment replacements, unplanned outages, and incidents associated with large property losses and injuries."

Readers can visit the US Chemical Safety Board website to view the animated video of Chevron Richmond Refinery, Richmond, California (www.csb.gov/chevron-refinery-fire).

Case Study 3

The Hydrocracker Explosion and Fire at BP Oil, Grangemouth Refinery, U.K., March 22, 1987

On Saturday March 22, 1987, the hydrocracker unit was being recommissioned following a routine shut down. During the recommissioning, a plant trip occurred. This was thought to be a spurious trip, and the plant operators started to bring up the unit to normal operating conditions. From then on until the time of the incident, the plant was held in standby condition with no fresh feed.

At 07.00 hours, the following morning there was an explosion and subsequent fire. The explosion could be heard and felt up to 30 km away. The explosion centered on a low pressure (LP) separator vessel, V306, which was fabricated from 18-mm steel plate and weighed 20 tonnes.

The investigation of the accident suggested that an air operated control valve on the high pressure (HP) separator had been opened and closed on manual control at least three times. Liquid level in the LP separator fell and the valve was opened. This action allowed the remaining liquid in the HP separator to drain away and for high pressure gas to break through into the LP separator. As the pressure relief on the LP separator had been designed for a fire relief case, not gas breakthrough, the vessel subsequently exploded.

Key Findings

The control valve did not close automatically because the low-level trip on the HP separator had been disconnected several years earlier. The operators did not trust the main level control reading and referred to a chart recorder for a backup level reading. There was an offset on this chart recorder which led them to assume that the level in the HP separator was normal.

Failures in Technical Measures

Operators did not trust the reading given by the float gauge in the HP separator and the nucleonic level indication was misread due to an offset on the chart recorder.

Maintenance procedure: sensors. The following aspects should be considered with respect to maintenance procedures:

- Human factors;
- Poorly skilled work force;

- Unconscious and conscious incompetence;
- Good maintainability principles;
- Knowledge of failure rate and maintainability; and
- Clear criteria for recognition of faults and marginal performance.

The following issues may contribute toward a major accident or hazard:

- Failure of safety critical equipment due to lack of maintenance;
- Human error during maintenance;
- Static or spark discharge during maintenance in an intrinsically safe zone;
- Incompetence of maintenance staff; and
- Poor communications between maintenance and production staff.

Control systems: sensors.

Operating procedures: provision of comprehensive written operating procedures.

The low-level trip on the HP separator had been disconnected several years earlier when a plant modification had been made.

Plant Modification/Change Procedures: HAZOP

The Chemical Industries Association (CIA) Safety Advisory Group outlines the following good practice related to plant modification/change procedures.

- Members of staff must be aware of the hazards associated with the work they carry out and be able to determine that the risks involved are acceptable.
- Risk assessment must be carried out to determine the possibility and consequence of the hazards being realized; if necessary, appropriate precautions must be taken to minimize the risk.
- All modifications—whether involving procedures, plant and equipment, people, or substances—should be subject to formal management procedures.

The procedure should draw reference to the following:

- Evidence from previous incidents—their cause and means of preventing them.
- The intrinsic link between process definition and the validity of the hazard evaluation.
- The options that are available in the design of safety measures.
- Preventative measures (process control, instrumentation, and so on.)
- Protective measures (containment, reactor venting, quenching, reaction inhibition).

Good industry practice requires that process and plant modifications should not be undertaken without having undertaken a safety, engineering, and technical review. This review should be traceable and identify changes proposed to the following factors:

- Process conditions
- Operating conditions
- Engineering methods
- Safety
- Environmental conditions
- Engineering hardware and design

A form of risk assessment should then identify what hazards have been created by the change that may affect plant or personnel safety and what action can be implemented to reduce or eliminate the risk. Additional hazards that may be introduced which need to be considered as fire, explosion, and loss of containment. Changes may affect other parts of the plant should be considered in undertaking hazard identification and risk assessments.

Factors that are crucial to the success and safe implementation of a plant modification procedure include the following:

- Corporate history.
- Communication between different departments.
- Recognition of authorized personnel.
- Accurate recording and monitoring of changes to plant and process.

HAZOP

Various stages of Hazard and Operability Study are generally undertaken. These are summarized in the table below.

HAZOP study	Description
1	Identify major hazards and check for availability of key hazard data.
2	Coarse HAZOP using flowsheet and block diagram.
3	Full HAZOP on frozen P & ID.
4	Check that all intended actions have been implemented, including hardware and software.
5	Pre-commissioning check including statutory requirement.
6	Safety audit after a few months of operation.

These are equally applicable to plant modifications as new plant. However, written procedures should be in place to determine what level of HAZOP should be applied. These procedures should take account of the following:

- The degree of hazard and extent of the change.
- The worst-case accident likely to arise because of the modification.
- Previous HAZOPs relating to the item being changed.
- The appointment of a competent person (excluding the HAZOP chairman) to decide the requirements for HAZOP.

A conventional hazard and operability study will identify potential hazards, but gives no likelihood of an incident occurring, or the loss suffered. The methodology has been developed to address likelihood and risk to assist in resource and priority allocation. Where hazards are not eliminated by actions placed during the HAZOP, hazard analysis should be employed to determine if the risk is acceptable (see Chapter 24 in Volume 4 of these series).

- Alarms/Trips/Interlocks: level measurement.
- The LP separator pressure relief had been sized for a fire case, not a gas breakthrough from the HP separator.
- Relief systems/vent systems: sizing of vents, relief valves.

Process plant can be subjected to excessive overpressure or underpressure due to the following:

- External fire.
- Process abnormality or maloperation.
- Equipment or service/utility failure.
- Changes in ambient conditions.
- Excess chemical reaction.

To achieve a more inherently safe design and to arrive at the most economical solution, overall consideration should always be given to the following:

- Can the overpressure or underpressure hazard be eliminated by changes in process or plant design?
- Can the overpressure or underpressure hazard be reduced by reducing inventories or changing process conditions?
- Can the overpressure or underpressure be contained by designing equipment to withstand maximum feasible pressure?
- Can alternative protection to a relief system be considered?
- Can the required relief system be minimized by mechanical or instrumented systems?

Relief systems for explosion relief are considered based on systems where pressure rise occurs over several seconds or longer, and there is no reaction front. In such cases, we can assume the following:

- Safety valves can open in time.
- Piping is adequately sized to provide pressure relief.
- Relief flow may be determined by steady state flow equations.
- Conditions are approximately uniform throughout each phase at any moment.
- Further pressure generation by reaction in the relief piping is negligible.

General principles applicable to relief systems include the following:

- In all cases, relief devices must be selected and located to minimize disturbance to plant and environment.
- Relief devices must not be isolated from equipment they protect while the equipment is in use.

The discharge from a relief device should pass to a safe location which may be as follows:

- A dump tank.
- Upstream in the process.
- A storage tank.
- A quench vessel or tower.
- A sewer.
- The atmosphere.
- A knockout drum.
- A scrubber.
- An incinerator.
- A flare stack.

Design basis and methodology of all relief stream packages must be documented and incorporated into plant modification and change procedures to ensure that relief stream invalidation does not occur.

There were no means of isolating the HP separator following an incident in the LP separator.

Isolation: emergency isolation.

Nomenclature: Distillation Process Performance

A, B thru K = Constants developed in original article
a, b, c = Correlation constants (distillation recoveries [166])
a = Activity of component
a_i = Activity of component, i
av or avg = Average
$a_{SRK}(T)$ = Attraction function in SRK equation 10.75, cm^6-bar/mol^2
B, C, D = Virial coefficients, Equation (19.11)
B = Bottoms product or waste, lb mols/hr, also = W
B_b = Moles of component, b, used as reference for volatility, after a given time of distillation
B_{bo} = Moles of component, b, used as reference for volatility, at start of distillation
B_i = Moles of component, i, after a given time of distillation
B_{io} = Moles of component, i, at start of distillation
B_{Ti} = Total mols of liquid in bottoms of still at time, T_1
B_{To} = Total mols liquid (not including any steam) in bottom of still at start time T_o (batch charge)
b = y intercept of operating line; or constant at fixed pressure for Winn's relative volatility
b_i = Moles of component, i, in bottoms
b_{SRK} = volume parameter, in SRK equation (19.76), cm^3/mol.
C = No. components present, phase rule; or no. components, or constant
C_{mi} = Factor in Colburn Minimum Reflux method, pinch conditions, stripping
C_{ni} = Factor in Colburn Minimum Reflux method, pinch conditions, rectifying
C = Specific heat, Btu/lb (°F)
C_{mi} = Moles of distillate or overhead product, lb mols/hr; or batch distillation, mols
d_i = Mole component, i, in distillate
E = Vaporization efficiency of steam distillation
E_G = Overall column efficiency
E_o = Overall tray efficiency
$E^{*}_{MV} = E_{oc}$ = Murphree point efficiency, fraction
$E_{MV}O$ = Murphree plate/tray efficiency, = E_M
F = Degrees of Freedom, phase rule; or, Feed, mols
F_{FR} = Factor for contribution of other feed flow to minimum reflux
F_L = Mole of liquid feed
F_v = Mole of vapor feed
$F_t = F + V_s$ = mole feed plus mols of non-condensable gases
$F_R = F_{SR,k}$ = Factor for contribution of side stream, k, flow to minimum reflux
F_{SR} = Factor for contribution of side stream flow, to minimum reflux
F_{FRJ} = Factor for contribution of feed, j, flow to minimum reflux
f = Fugacity at a specific condition
fυ = Fugacity at reference standard condition
f_i = Feed composition, i; or, = total mols of component, i, in distillate and bottoms
G = Boil-up rate, mols/h
H = Total enthalpy, above reference datum, of vapor mixture at tray or specified conditions, Btu/lb mol, or Btu/lb
$H' = H_{ij}$ = Henry's Law constant, lb mols/(cu ft) (atm)
H_n = Total molal enthalpy of vapor at conditions of tray, n, entering tray; $H_n = \Sigma H_{ni} (y_{ni})$
H_s = Total enthalpy of steam, Btu/lb mol, or Btu/lb
HK = Heavy key component in volatile mixture
h = Enthalpy of liquid mixture or pure compound at tray conditions of temperature and pressure, or specified point or condition, Btu/lb mol, or Btu/lb

h_n = Total molal enthalpy of liquid at conditions of tray, n; $h'_n = h_{ni}(x_{ni})$
h_D = Molal enthalpy of product or total liquid enthalpy above reference datum for sum of all contributing percentages of individual components
h_{n+1} = Molal enthalpy of liquid leaving plate n + 1
h_a = Total molal enthalpy of liquid at conditions of tray, n; $h_n = \Sigma h_{ni} (x_{ni})$
K = Equilibrium constant for a particular system (= y/x)
K′ = Equilibrium constant for least volatile component, K′ = y/x
K_i = Equilibrium distribution coefficient for component, i, in system
k = Experimentally determined Henry's Law constant, also can be K
kpa = Metric pressure
L = Liquid flow rate return to tower as reflux, lb mols/h, mols component in liquid phase; or L_1, L_2 = Latent heat of vaporization; or volumetric flow rate for incoming contaminated water (stripping VOC with air); or mols liquid produced from F per unit time, leaving flash zone, or moles liquid per unit time, rectifying section.
$\bar{L}$ = moles liquid from feed plate per unit time.
L_o = moles reflux liquid returned to plate 1 per unit time.
L_r = liquid flow rate down rectifying section of distillation tower, lb mol/h
L_s = liquid flow rate down stripping section of distillation tower, lb mol/h
LK = Light key component in volatile mixture
L/V = Internal reflux ratio
L/D = Actual external reflux ratio
$(L/D)_{min}$ = Minimum external reflux ratio
M = Molecular weight of compound
M_s = Total mole steam required
m = Number of side streams above feed, n
N = Number of theoretical trays in distillation tower (not including reboiler) at operating finite reflux. For partial condenser system N includes condenser; or number theoretical trays or transfer units for a packed tower (VOC calculations)
N_B = Number of trays from tray, m, to bottom tray, but not including still or reboiler
N_{min} = Minimum number of theoretical trays in distillation tower (not including reboiler) at total or infinite reflux. For partial condenser system, N_{min} includes condenser; also, minimum value of N
N_n = Number of theoretical trays above feed, or reference plate, n, but not including n
N_m = Number of theoretical trays before feed tray
N_{im} = Moles of immiscible liquid
N_o = Moles of non-volatile material present; or, number of theoretical trays/stages in column only, not reboiler or condenser.
N_s = Moles of steam
n = Number of theoretical trays in rectifying section or number of components, or minimum number of equilibrium trays; number of plates between any two points in the column under consideration.
n_f = Number of feeds
n_s = Number of side streams
P = Pressure, atmospheres; or vapor pressure of component, atm.; or
P = number of phases; or P = for batch operations, percentage draw-off
P_i = Vapor pressure of each component
P_s = Vapor pressure of steam, absolute
P_b = Vapor pressure of reference more volatile component, b
p = Pi = Partial pressure of one compound in liquid, absolute units, or ratio rps/rpr; also, P_i = partial pressure of solute (Henry's Law)
p = Total pressure of system = π
p′ = Number of trays below feed where introduction of light components should begin, Akers–Wade calculation method

P_i^* = Vapor pressure component, i, in pure state at temperature

P_{ii}^* = Similar to above by analogy

p” = Number of trays above feed where introduction of heavy components should begin. Akers–Wade calculation

p_{im} = Pure component vapor pressure of immiscible liquid, mm Hg

p_s = Partial pressure of steam, mm Hg

Q_B = Net heat in through reboiler duty, Btu/hr; or heat added in still or bottoms

Q_c = Net heat out of overhead condenser, Btu/hr, = w cp $(t_i - t_o)$

Q = qF = Thermal condition of feed, approximately amount of heat to vaporize one mol of feed at feed tray conditions divided by latent heat qf vaporization of feed

qB = reboiler duty, Btu/h

qD = condenser duty, Btu/h

q_v = Thermal condition of side stream (s)

R = Reflux ratio = External reflux ratio for a given separation, = L/D, L = liquid rectifying column

R = Actlial reflux ratio, O/D

R_m = Minimum reflux ratio, O/D

R′ = Pseudo minimum reflux ratio

R_{min} = Minimum external reflux ratio for a given separation

R_F = Feed component of minimum reflux

$R_{F,n}$ = Feed component of minimum reflux for feed, n

R_{OF} = Other feed components of minimum reflux

R_s = Side stream component of minimum reflux

r_{ps} = Ratio of light to heavy keys, stripping pinch

r_{pr} = Ratio of light to heavy keys, rectifying pinch

r_f = Ratio mol fraction light key to heavy key in feed

S_n = Theoretical stages at minimum reflux

S_M = Minimum theoretical stages at total reflux from bottoms composition through overhead product composition, including reboiler and any partial condenser (if used); or minimum stripping fat for at minimum flow rate of air.

S_k = Flow rate of side stream, k, mol/h

S_o = Theoretical stages at a finite operating reflux; or batch, mols originally charged to kettle

S_r = Theoretical stages in total rectifying section, including partial condenser, if used

S_s = Theoretical stages in total stripping section, including reboiler

S_t = Theoretical trays/stages at actual reflux, L/D, including reboiler and total condenser

S_{opt} = Optimum stripping factor

$(SR)_i$ = Separation factor

s = Pounds (or moles) steam per pound (or mol) of bottoms; or flow rate of side stream, mol/h

T = Temperature, °Abs R

t_B = Bottoms temperature, °F

t_i = Temperature in, °F

t_o = Temperature out, °F; or overhead temperature, °F

$V = V_t$ = Total vapor leaving flash zone/unit time at specific temperature and pressure; or total overhead vapor from tower, mol/h; or mols of component in vapor phase; or volumetric flow rate for incoming fresh air

V = Quantity of vapor, mols

V_C = moles of vapor condensed necessary to heat reflux liquid to t.

V_F = moles vapor per unit time rising from feed plate.

V_r = Vapor flow rate up rectifying section of tower, lb mol/h

V_s = Vapor flow rate up stripping section of tower, lb mol/h; or mole non- condensable gases entering with feed, F, and leaving with vapor, V/time

V_{min} = Minimum fresh air flow based on slope of operating line, L/V, on x-y diagram

v = Vapor flow rate, mols/h; or molar volume
W = Bottoms product, or still bottoms, or kettle bottoms, mols; also see B; or mols/hr bottoms product; or mols of residue or bottoms/unit time (Ponchon heat balance)
$x_i = x$ = Mole fraction of component in liquid phase; or mol fraction solute in solution (Henry’s Law)
x’ = Mol fraction of least volatile component
x'_1 = $x_l - k$
x'_p = $x - k$
x = Mole fraction more volatile component in liquid
x_1 = Mole fraction of component, i, in liquid mixtures as may be feed distillate or bottoms, B_T, at any time, T_l; or mol fraction more volatile in vapor entering column at any time (or in distillate)
x_{it} = Mole fraction liquid at intersection of operating lines at minimum reflux, Scheibel–Montross equation
x_{hf} = Mole fraction heavy key in feed
x_n = Pinch composition and light component mol fraction
x_N = Mole fraction VOC component in the stripped water exiting, usually targeted at meeting environmental regulations
x_{iD} = Mole fraction light key component in overhead product; or, any light component (Colburn)
x_{iB} = Mole fraction light key component in keys in original charge
x_{io} = Mole fraction light key in overhead expressed as fraction of total keys in overhead
x_{1B} = Mole fraction most volatile component in bottoms
x_{hD} = Overhead composition of heavy key component, mole fraction
x_{hn} = Pinch composition of heavy key component, mole fraction
x_l = Mole fraction of component in liquid phase; or mol fraction more volatile component in vapor entering column at any time
x_F = Mole fraction more volatile component in feed
x_D = Mole fraction more volatile component in final distillate = mol fraction in distillate leaving condenser at time θ
x_p = Mole fraction of more volatile component in liquid leaving column at any time
x_L = Mole fraction of feed as liquid, Scheibel–Montross
x_{lo} = Mole fraction light key in overhead expressed as fraction of total keys in overhead, Scheibel–Montross equation
x_m = Tray liquid mol fraction for start of calculations (most volatile component)
x_o = Mole fraction of component, i, in bottoms B_{To} at start time, T_o; or VOC mol fraction
xv = Mole fraction of feed as vapor, Scheibel–Montross equation
$y = y_i$ = Mole fraction of component in vapor phase, as may be feed, distillate, or bottoms; or Henry’s Law, y_i = mol fraction solute in vapor
y_i = Mole fraction VOC component in the exiting VOC contaminated air
y’ = Mole fraction of least volatile component
y^* = Equilibrium value corresponding to x_i
y_n = Average light key mol fraction vapor leaving plate, n
y_{n-1} = Average light key mol fraction vapor entering plate, n + 1
y_{n-1} = Mole fraction VOC component in the incoming fresh air (equals zero for fresh air)
y_j = Mole fraction solvent component in vapor
y_s = Mole fraction steam in vapor
Y_{iB} = Percent recovery of, i, in the bottoms
Y_{iD} = Percent recovery of, i, in the distillate
Z = Compressibility factor
$z_{i,F}$ = Mole fraction component, i, in feed
$z_{i,Fj}$ = Mole fraction component, i, in feed, j
$z_{i,S}$ = Mole fraction component, i, in side stream
$z_{i,Sk}$ = Mole fraction component, i, in side stream, k

Greek Symbols

α, α_1 = Relative volatility of light key to heavy key component, or any component related to the heavy key component, except Equation (19.65), α_i is based on heavy key
α_{avg} = Average relative volatility between top and bottom sections of distillation tower/column
α_j = Relative volatility of more volatile to each of other components (steam distillation)
αi = Relative volatility of component, i
α_H = Relative volatility of components heavier than heavy key at feed tray temperature
α_i = Relative volatility of more volatile to each of other components
α_L = Relative volatility of components lighter than light key at feed tray temperature
β = Constant of fixed pressure in Wino's relative volatility, Equation 19.43
θ = Time from start of distillation to fill receiver, or value of relative volatility (Underwood Parameter) to satisfy Underwood Algebraic Method
μ = Viscosity, centipoise
υ = Activity, coefficient
π = total system pressure, absolute; atm, mm Hg, psia
π = 3.14159
Σ = Sum
ψ = First derivative function
ψ' = Second derivative function
ω_j = Function in Underwood's Algebraic method for minimum reflux ratio
Ω = Fugacity, coefficient
ϕ = No. phases from phase rule

Subscripts

a, b, c, etc. = Specific components in a system or mixture
Avg, Av = Average
B = Any consistent component in bottoms product
B = b = Bottoms
b = Exponent in Winn's relative volatility equation
D = Any consistent component in condensed overhead product or distillate
eff = Effective
F = Feed
F_j = Feed, j
F_n = Intermediate feed, Scheibel–Montross method
F_L = FH = All mol fractions lighter than light key in feed, Scheibel–Montross method
FHK = Heavy key in feed
FLK = Light key in feed
HK = h = h_k = Heavy key component
H = Components heavier than heavy key
h = Heavy, or heavy or high boiling component in mixture; also heavy key component
i = Any component identified by subscripts 1, 2, 3, etc., or by a, b, c, etc.; or initial condition, i
im = Immiscible liquid
j = Specific components in a system or mixture
1 = 1k = Light key component; or light or low boiling component in mixture
1h = Refers to light component referenced to heavy component
LK = Light key component
L = Liquid, Scheibel–Montross method only; or components lighter than light key

M = min = Minimum
m = No. trays in stripping section; or tray number
n = No. trays in rectifying section; or tray number
o = Initial conditions; or i; or operating condition
p_r = Pinch condition in rectifying section
p_s = Pinch condition in stripping section
P = For packed towers
w = Relates to bottoms or pot liquor, or kettle bottoms
r = Rectifying section; or component to which all the relative volatilities are referred
s = Steam, or stripping section of column
t = Top, or total
T = For tray towers
v = Vapor
1 = Initial, steam distillation
2 = Remaining, steam distillation
1,2,3, etc. = Tray numbers; or specific components in a system or mixture
(‘) = Superscript, heavy key component or least volatile

References

1. Akers, W. W. and D. E. Wade, “New Plate-to-Plate Method,” *Pet. Ref.* V. 36, p. 199, (1954).
2. American Institute of Chemical Engineers, “Bubble Tray Design Manual, Prediction of Fractionation Efficiency,” Amer. Inst. Chem. Engrs., (1958).
3. Biggers, M. W., private communication.
4. Bogart, M. J. P., “The Design of Equipment for Fractional Batch Distillations,” *Trans.* A.I.Ch.E. V. 33, p. 139, (1937).
5. Bolles, W. L., “Optimum Bubble-Cap Tray Design,” *Pet. Processing;* Feb. through May, 1956.
6. Boston and Sullivan, *Canadian Jour. of Chem. Engr:* V. 50, Oct., (1972).
7. Broaddus, J. E., A. J. Moose, R. L. Huntington, “How to Drain Bubble Cap Columns,”. *Pet. Ref.,* Feb., (1955).
8. Brown, G. G. and Associates, *Unit Operations,* 4th Ed. John Wiley and Sons, New York, N.Y., (1953).
9. Brown, G. G. and H. Z. Martin; “An Empirical Relationship Between Reflux Ratio and the Number of Equilibrium Plates in Fractionating Columns,” *Trans.* A.I.Ch.E. V. 38, No. 5, (1939).
10. Chueh, P. L. and J. M. Prausnitz, I.&E.C. Fundamentals, Amer. Chem. Society, V. 6, p. 492, (1967).
11. Cicalese, J. J., J. J. Davis, P. J. Harrington, G. S. Houghland, A. J. L. Hutchinson, and T. J. Walsh, *Pet. Ref.* V. 26, May, p.127, (1947).
12. Colburn, A. P., “Calculation of Minimum Reflux Ratio in Distillation of Multicomponent Mixtures,” *Trans.* A.I.Ch.E. V. 37, p. 805, (1941).
13. Dauphine, T. C., “Pressure Drops in Bubble Trays,” Sc. D., Thesis, Mass. Inst. Technology, (1939).
14. Davies, J. A., “Bubble Tray Hydraulics,” *Ind. Eng. Chem.* V. 39, p. 774, Amer. Chem. Society, (1947).
15. Davies, J. A., “Bubble Trays-Design and Layout,” *Pet. Ref.* V. 29, 93, p. 121, (1950).
16. Drickhamer, H. G. and J. B. Bradford, “Overall Plate Efficiency of Commercial Hydrocarbon Fractionating Columns,” Trans. A.I.Ch.E. V. 39, p. 319 (1943).
17. Edmister, W. C., “Design for Hydrocarbon Absorption and Stripping,” *Ind. Eng. Chem.* V. 35, p. 837 (1943).
18. Edmister, W. C., “Hydrocarbon Absorption and Fractionation Process Design Methods,” *Pet. Engr.* May, 1947-March, 1949 and, “Absorption and Stripping-Factor Functions for Distillation Calculations by Manual and Digital-Computer Methods.” *A.I.Ch.E. Journal,* V. 3, No. 2 p. 165, (1957).
19. Eduljee, H. E., “Entrainment From Bubble-Cap Distillation Plates,” *British Chem. Engr.,* p. 474, Sept., (1958).
20. Ewell, R. H., J. M. Harrison, and Lloyd Berg, “Hydrocarbon Azeotropes,” *Pet. Engr.,* (installment, Oct., Nov., Dec., (1944)).
21. Faassen, J. W., “Chart for Distillation of Binary Mixtures,” *Ind. Eng. Chem.,* V. 36, p. 248, (1944).
22. Gautreaux, M. F., H. E. O’Connell, “Effect of Length of Liquid Path on Plate Efficiency,” *Chem. Eng. Prog.* V. 51, p. 232, (1955).
23. Gilliland, E. R., “Multicomponent Rectification,” *Ind. Eng. Chem.* V. 32, pp. 1101 and 1220, (1940).

24. Good, A. J., M. H. Hutchinson, W. C. Rousseau, "Liquid Capacity of Bubble Cap Plates," *Ind. Eng. Chem.,* V. 34, p. 1445, (1942).
25. Holland, C. D., Advanced Distillation Course in Extension, Texas A&M College, (1954).
26. Holland, C. D., *Multicomponent Distillation,* Prentice-Hall, (1963).
27. Holland, C. D., *Unsteady State Processes with Applications in Multicomponent Distillation,* Prentice-Hall.
28. Horsley, L. H., "Azeotropic Data," *Advances in Chemistry* Series, American Chemical Society, Washington, D.C.
29. Horton, G., W, B. Franklin, "Calculation of Absorber Performance and Design," *Ind. Eng. Chem.* V, 32, p. 1384, (1940).
30. Huang, Chen-Jung, and J. R. Hodson, "Perforated Trays Designed This Way," *Pet. Ref.* V. 37, p. 104, (1958).
31. Hughmark, G. A., and H. E. O'Connell, "Design of Perforated Plate Fractionating Towers," *Chem. Eng. Prog.* V. 53, p. 127-M, (1957).
32. Hull, R. J. and K. Raymond, "How To Design and Evaluate Absorbers," *Oil and Gas Jour.,* No. 9, 1953 through Mar. 15, (1954).
33. Hunt, C. D. A., D. N. Hanson and C. R. Wilke, "Capacity Factors in the Performance of Perforated Plate Columns," *A.I.Ch.E. Jour.,* V. 1, p. 441, (1955).
34. Hutchinson, A. J. L., "A System of Calculations for Light Hydrocarbons," *Pet. Ref.* Oct. 1950-April, (1951).
35. Jennings, B. H. and F. P. Shannon, "Aqua-Ammonia Tables," Lehigh University, Part 1, Science and Technology. Series No.1, Bethlehem, Pa.
36. Jones, J. B. and C. Pyle, "Relative Performance of Sieve and Bubble Cap Plates," *Chem. Eng. Progress,* V. 51, p. 24, (1955).
37. Kelly, R. G., *Oil and Gas Journal,* April 18, p. 128, (1955).
38. Kemp, H. S. and C. Pyle, "Hydraulic Gradient Across Various Bubble Cap Plates," *Chem. Eng. Prog. V.*45, p. 435, (1949).
39. Klein, J. H., D. Sc. Thesis, Mass. Inst. Technology, (1950).
40. Kremser, A., "Theoretical Analysis of Absorption Process," *Nat. Pet. News,* V. 22, p. 48, (1930).
41. Lee, D. C., Jr., "Sieve Trays," *Chem. Eng.* p. 179, May, (1954).
42. Leibson, I., R. E. Kelley and L. A. Bullington, "How to Design Perforated Trays," *Pet. Ref., V.* 36, p. 127, (1957).
43. Martin, G. Q., "Guide To Predicting Azeotropes," *Hydrocarbon Processing,* No. l, p. 241, (1975).
44. May, J. A. and J. C. Frank, "Compensation for Hydraulic Gradient in Large Fractionator," *Chem. Eng. Prog., V.*51, p. 189, (1955).
45. Mayfield, F. D., W. L. Church, Jr., A. C. Green, D. C. Lee, Jr. and R. W. Rasmussen, "Perforated-Plate Distillation Columns," *Ind. Eng. Chem.* V. 44, p. 2238, (1952).
46. Munk, Paul, "Design of Bubble Cap Trays," *Pet. Ref.* V. 34, p. 104, (1955).
47. Myers, H. S., "A Versatile Fractionating Column," *Ind. Eng. Chem.* V. 50, p. 1671, (1958).
48. *Natural Gasoline Supply Men's Association, Engineering Data Book,* 7th. Ed. 1957, Tulsa, Oklahoma.
49. O'Connell, H. E., "Plate Efficiency of Fractionating Columns and Absorbers," *Trans.* A.I.Ch.E., V. 42. p. 741, (1946).
50. Orye, R. V. and J. M. Prausnitz, *Ind. Eng. Chem. 57,* 5, p. 19, (1965).
51. Palmer, D. A., "Predicting Equilibrium Relationships for Maverick Mixtures," *Chem. Eng., June* 9, p. 80, (1975).
52. Prausnitz, J. M. and P. L. Cheuh, "*Computer Calculations for High Pressure Vapor-Liquid Equilibrium,*" Prentice-Hall Inc., (1968).
53. Prausnitz, J. M., C. A. Eckert, R. V. Orye and J. P. O'Connell, "*Computer Calculations for Multicomponent V-L Equilibria,*" Prentice-Hall, (1967).
54. Prausnitz, J. M., "*Molecular Thermodynamics of Fluid Phase Equilibria,*" Prentice-Hall, (1969).
55. Redlich, O. and A, T. Kister, *Ind. Eng. Chem., V.* 40, p. 345, (1948).
56. Redlich, O., and J. N. S. Kwong, *Chem. Rev.* V. 44, p, 233, (1949).
57. Redlich, O., T. Kister, and C. E. Turnquist, *Chem. Engr. Progr. Sym. Ser.* 48, 2, p. 49, (1952).
58. Renon, H. and J. M. Prausnitz, *A.I.Ch.E. Jour.* V. 14, p.135, (1968).
59. Robinson, C. S. R. and E. R. Gilliland, *Elements of Fractional Distillation,* McGraw-Hill, 4th Ed., (1950).
60. Rogers, M. C. and E. W. Thiele, "Pressure Drop in Bubble-Cap Columns," *Ind. Eng. Chem.* V. 26, p. 524, (1934).
61. Scheibel, E. G. and C. F. Montross, "Empirical Equation for Theoretical Minimum Reflux," *Ind. Eng. Chem. V.* 38, p. 268, (1946).
62. Sherwood, T. K., *Absorption and Extraction,* McGraw-Hill Book Co., Inc., New York, N .Y., (1937).
63. Shiras, R. N., D. N. Hansell, C. H; Gibson, "Calculation of Minimum Reflux in Distillation Columns," *Ind. Eng. Chem., V. 42*. p. 871, (1950).
64. Simkin, D. J., C. P. Strand and R. B. Olney, "Entrainment from Bubble Caps," *Chem. Eng. Prog.,* V. 50, p. 565, (1954).
65. Smith, B. D., *Design of Equilibrium Stage Processes,* McGraw-Hill, (1963).
66. Smoker, E. H., "Nomographs for Minimum Reflux Ratio and Theoretical Plates for Separation of Binary Mixtures," *Ind. Eng. Chem.,* V. 34, p. 509, (1942).

67. Souders, M., Jr., G. G. Brown, "Fundamental Design of Absorbing and Stripping Columns for Complex Vapors," *Ind. Eng. Chem.,* V. 24, p. 519, (1932).
68. Souders, M., Jr., G. G. Brown, "Design of Fractionating Columns," *Ind. Eng. Chem.,* V. 26, p. 98, (1934).
69. Sutherland, S., Jr., "Characteristics of Countercurrent Vapor Liquid Flow at a Perforated Plate," M. S. Thesis, Texas A&M College, Jan., (1958).
70. Teller, A. J., "Binary Distillation," *Chem. Eng.,* p. 168, Sept., (1954).
71. Umholtz, C. L. and M. Van Winkle, "Effect of Hole Free Area, Hole Diameter, Hole Spacing Weir Height, and Downcomer Area," *Pet. Ref.,* V. 34, p. 114, (1955).
72. Underwood, A. J. V., "Fractional Distillation of Multicomponent Mixtures," *Chem. Eng. Prog., V.* 44, p. 603, (1948).
73. Underwood, A. J; V., *Trans. Inst.,* Ch. E. (London) V. 10 p. 112, (1932).
74. Van Winkle, M., *Distillation,* McGraw-Hill, Inc., (1967).
75. Van Winkle, M., "Multicomponent Distillation," *Oil and Gas Journal,* p. 182, Mar. 23, (1953).
76. Wang, J. C. and G. E. Henke, "Tridiagonal Matrix for Distillation," *Hydrocarbon Processing,* V. 45, No.8, p. 155, (1966).
77. Wilson, G. *M., J. Am. Chem. Soc.,* V. 86, p. 127, (1964).
78. Zenz, F. A., "Calculate Capacity of Perforated Plates," *Pet. Ref., V.*.33, p. 99, (1954).
79. Gas Processors Suppliers Association; *Engineering Data Book,* V 1 and 2, 3rd, Rev., 12th. Ed., (2004), produced in cooperation with the Gas Processors Association.
80. DePriester; C. L., The American Institute of Chemical Engineers; *Chem. Eng. Prog. Sym. Ser.,* V. 49, No.7, p. 1, (1953).
81. Nelson, W. L., *"Petroleum Refinery Engineering,"* 1st. Ed. McGraw-Hill Co., Inc., p. 244, (1936).
82. Carroll, J. J., "What is Henry's Law?" *Chem. Eng. Prog* V. 87, p. 48 Sept., (1991).
83. Carroll, J. J., "Use Henry's Law for Multicomponent Mixtures," *Chem. Eng. Prog.,* V. 88, No.8, p. 53, (1992).
84. Eduljee, H. E., "Equations Replace Gilliland Plot," *Hydro. Proc.* V. 54, No. 9, p. 120, (1975).
85. Murphree, E. V., *Ind. Eng. Chem.* V. 17, p. 747, (1925).
86. MacFarland, S. A., P. M., Sigmund, and M.Van Winkle, "Predict Distillation Efficiency," *Hydro. Proc.* V. 51, No. 7, p 111, (1972).
87. Biddulph, M. W., "When Distillation Can Be Unstable," *Hydro. Proc.* V. 54, No.9, p. 123, (1975).
88. Fair, J, R., "Tray Hydraulics-Perforated Trays," Chap. 15 in *Design of Equilibrium Stage Processes,* Smith, B. D., McGraw-Hill, New York, p. 552, (1963).
89. Leva, M., "Film Tray Equipment for Vacuum Distillation," *Chem. Eng. Prog.,* V. 67, No.3, p. 65, (1971).
90. Biddulph, M. W., "Tray Efficiency Is Not Constant," *Hydro. Proc.,* p, 145, Oct., (1977).
91. Wichterle. I., R., Kobayashi. and P. S. Chappelear, "Caution! Pinch Point in Y-X Curve," *Hydro. Proc., p. 223, Nov.,* (1971).
92. Wagle, M. P., "Estimate Relative Volatility Quickly," *Chem. Eng.,* V. 92, No. 9, p. 83, (1985).
93. Reid, R. C., J. M, Prausnitz, and T. K. Sherwood, *"The Properties of Gases and Liquids,"*, 3rd. Ed., McGraw-Hill, New York, p. 214, (1977).
94. Kister, H. Z., *"Distillation Design,"* McGraw-Hi1l, Inc. (1992).
95. Kister, H. Z., "Complex Binary Distillation," *Chem. Eng.,* V. 92, No.2, p. 97, (1985).
96. Chou, S-M. and C. L. Yaws, "Minimum Reflux for Complex Distillations," *Chem. Eng., V.* 95, No. 6, p. 79, (1988).
97. McCormick, J. E., "A Correlation for Distillation Stages and Reflux," *Chem. Eng., V.* 95, No.13, p 73, (1988).
98. Venkateswara, Rao K., and A. Raviprasad, "Quickly Determine Multicomponent Minimum Reflux Rates," *Chem. Eng.,* V. 14, p. 137, (1987).
99. Winn, F. W., "New Relative Volatility Method for Distillation Calculations," *Pet. Refiner* (now *Hydro. Proc.)* V. 37, No.5, p. 216, (1958).
100. Frank, O., "Shortcuts for Distillation Design," *Chem. Eng.,* p. 109, Mar. 14, (1977).
101. Kessler, D. P. and P. C. Wankat, "Correlations for Column Parameters," *Chem. Eng.,* V. 95, No. 13. p. 71, (1988).
102. Tsuo, Fu-Ming, C. L. Yaws and J-S. Cheng, "Minimum Reflux for Sidestream Columns," *Chem. Eng.*, p. 49, July. 21, (1986).
103. Chou, S-M. and C. L. Yaws, "Reflux for Multifeed Distillation," *Hydro; Proc.*, p. 41, Dec., (1986).
104. Scheiman, A. D., "Find Minimum Reflux by Heat Balance," *Hydro. Proc.*, p. 187, Sept., (1969).
105. Lessi, A., "New Way to Figure Minimum Reflux," *Hydro. Proc.,* V. 45, No.3, p. 173, (1966).
106. Hengstebeck, R. J., *"Distillation-Principles and Design Procedures,"* Reinhold. Pub. Co. New York, p. 120, (1961).
107. Eduljee, H. E., "Easy Way to Remember Reflux vs. Trays," *Hydro. Proc.,* V. 42, No. 3, p 185, (1963).
108. Maas, J. H., "Optimum-Feed Stage Locations in Multicomponent Distillations," *Chem. Eng.,* Apr. 16, p. 96, (1973).
109. Dechman, D. A., "Correcting the McCabe-Thiele Method for Unequal Molal Overflow," *Chem. Eng.*, p. 79, Dec. 21, (1964).
110. Torres-Marchal, C., "Graphical Design for Ternary Distillation Systems," *Chem. Eng.,* V. 88. No.21, p. 134, (1981).
111. Torres-Marchal, C., "Calculating Vapor-Liquid Equilibria for Ternary Systems," *Chem. Eng., V.* 88, No.21, p. 141, (1981).
112. Maddox, R. N., "Calculations for Multicomponent Distillation," *Chem. Eng.,* Dec, 11, p. 127, (1961).

113. Erbar, R. C., R. S. Joyner, and R. N. Maddox, "For Multi-component Columns, How to Calculate Minimum Reflux," *Petro/Chem.* Engr., p. C-19, Mar., (1961).
114. Lyster,W. N., S. L. Sullivan, D. S. Billingsley, and C. D. Holland, "Digital Computer Used to Figure Distillation This New Way," *Pet. Ref.* (now, Hydro. Proc.), p. 221, June, (1959).
115. Maddox, R. N., and J. H. Erbar, "Programming Plate to Plate Distillation Calculations," *Refining Engr., p. C-35,* Sept., (1959).
116. Maddox, R. N., and W. A., Jr., Fling, "Try the Pseudo-K Method for Short-cut Multicomponent Distillation Columns," *Petro/Chem. Engr., p. C-37,* Mar., (1961).
117. Kister, H. Z., "*Distillation operation,*" McGraw-Hill, Inc., (1990).
118. Forman, E. R., "Control Systems for Distillation," *Chem. Eng.,* p. 213, Nov.8, (1965).
119. Hoffman, H. L., "HP/PR Special Report, Automatic Control for Distillation," *Hydro. Proc.* and *Pet. Ref.,* V. 42, p. 107, No. 2, (1963).
120. Shinskey, F. G., "*Process-Control Systems,*" McGraw-Hill Book Co., 2nd. Ed., (1979).
121. Trigueros, D., Coronado-Velasco, C. and A. Gornez-Munoz, "Synthesize Simple Distillation the Thermodynamic Way," *Chem. Eng.,* V. 96, No.8, p. 129, (1989).
122. Mapstone, G. E., "Reflux versus Tray's By Chart," Hydroc. Proc., V. 47, No.5, p. 169, (1968).
123. Zanker, A, "Nomograph Replaces Gilliland Plot," Hydroc. Proc., p. 263, May, (1977).
124. Yaws, C. L., P. M. Patel; F. H. Pitts, and C. S. Fang, "Estimate Multicomponent Recovery," *Hydro. Proc.,* V. 58, No.2, p. 99, (1979).
125. Hengstebeck, R.J. *Trans. A.I.Ch.E.,* V. 42, p. 309, (1946).
126. Geddes, R. L., *A.I.Ch.E. Journ.,* V. 4, p. 389, (1958).
127. Ellerbe, R .W., "Steam-Distillation Basics," *Chem. Eng.,* p. 105, Mar. 4, (1974).
128. Ellerbe, R. W., Section 1.4 in Schweitzer, P. A. (editor), *Handbook of Separation Techniques for Chemical Engineers,* McGraw-Hill Book Co., (1979).
129. Treybal, R. E., "A Simple Method for Batch Distillation," *Chem. Eng.,* p. 95, Oct. 5, (1970).
130. Ellerbe, R.W., "Batch Distillation Basics," *Chem. Eng.,* p. 110, May 28, (1973).
131. Schweitzer, P. A. "*Handbook of Separation for Chemical Engineers,*" McGraw-Hill Book Co., (1979).
132. Perry, R. H., and Green, D. *"Perry's Chemical Engineers Handbook,"* 6th. Ed. McGraw-Hill Book Co., (1984).
133. Block, B., "Batch Distillation of Binary Mixtures Provides Versatile Process Operations," *Chem. Eng.,* p. 87, Feb.6, (1961).
134. Block, B., "Control of Batch Distillations," *Chem. Eng.,* p. 147, Jan. 16, (1967).
135. Brasens, J. R., "For Quicker Distillation Estimates," *Hydro. Proc., p. 102,* Oct., (1969).
136. Liddle, C. J., "Improved Short-cut Method for Distillation Calculations," *Chem. Eng.*, p. 137, Oct. 21 (1968).
137. Hengstebeck, R. J., "An Improved Shortcut for Calculating Difficult Multicomponent Distillations," *Chem. Eng.*, p. 115, Jan. 13 (1969).
138. Guy, J. L., "Modeling Batch Distillation in Multitray Columns," *Chem. Eng.*, p. 99, Jan. 10, (1983).
139. Van Winkle, M. and W. G. Todd, "Optimum Fractionation Design by Simple Graphical Methods," *Chem. Eng.,* p. 136, Sept 20, (1971).
140. Koppel, P. M., "Fast Way to Solve Problems for Batch Distillation," *Chem. Eng.,* p. 109, Oct. 16, (1972).
141. Yaws, C. L., C-S. Fang, and P. M. Patel, "Estimating Recoveries in Multicomponent Distillation," *Chem. Eng.*, p. 101, Jan. 29, (1979).
142. Kumana, J. D., "Run Batch Distillation Processes with Spreadsheet Software," *Chem. Eng. Prog.,* p. 53, Dec., (1990).
143. Li, K. Y. and K. J. Hsiao, "How to Optimize an Air Stripper," *Chem. Eng.,* V. 98, No.7, p. 114, (1991).
144. Salmon, R., "New Method for V /L Flash Calculations," *Pet. Ref.,* (now *Hydro. Proc.)* V. 36, No.12, p. 133, (1957).
145. Lockhart, F. J., and R. J., McHenry, "Figure Flash Equilibria Quicker This Way," *Pet. Ref.,* V. 37, No.3, p. 209, (1958).
146. Gallagher, J. L., "New Short-cut for Flash Calculations," *Hydro. Proc.,* V. 42, No. 2, p. 157, (1963).
147. Holland, C. D. and P. R., Davidson, "Simplify Flash Distillation Calculations," *Pet. Ref .,* V. 36, No.3, p. 183, (1957).
148. Deam, J. R. and R. N., Maddox, "How to Figure Three-Phase Flash," *Hydro. Proc.,* V. 51, No.10, p. 163, (1969).
149. Henley, E. J. and L. D., Seader, *"Equilibrium-Stage, Separation Operations in Chemical Engineering,"* John Wiley and Sons, Inc., (1981).
150. Eckles, A., P., Benz, and S., Fine, "When to Use High Vacuum Distillation," *Chem. Eng* V. 98, No. 5, p. 201, (1991).
151. Elsby, K., *et al.*, "Packing Performance in Vacuum Distillation," *A.I.Ch.E., Symposium, Series No.104 Distillation and Adsorption,* pp. 143-148, (1987).
152. *"Vacuum Technology for Chemical Engineering,"* Leybold, AG., p. 11.
153. Holland, C. D., S. E., Gallun, and M. J., Lockett, "Modelling Azeotropic and Extractive Distillation," *Chem. Eng.,* p. 185, Mar. 23, (1981).

154. Luyben, W. L., "Azeotropic Tower Design by Graph," *Hydro.* Proc., p. 109, Jan., (1973).
155. Stichlmair, J. J., J. R., Fair, and J. L., Bravo, "Separation of Azeotropic Mixtures via Enhanced Distillation," *Chem. Eng. Prog.,* V. 85, No.1, p. 63, (1989).
156. Gerster, J. A., "Azeotropic and Extractive Distillation," *Chem. Eng. Prog.,* V. 65, No. 9 p. 43, (1969).
157. Holland, C. D., *et al.*, "Solve More Distillation Problems," 13 Parts; *Hydrocarbon Proceessing,* Part 1, V. 53, No. 7, (1974); Part 2, V. 53, No. 11, (1974); Part 3, V. 54, No. 1 (1975); Part 4, V. 54, No. 7, (1975); Part 5, V. 55, No. 1, (1976); Part 6, V. 55, No. 6, (1976); Part 7, V. 56 No. 5, (1977); Part 8, V. 56, No. 6, (1977); Part 9, V. 59, No. 4., (1980); Part 10, V. 59, No. 7, (1980); Part 11, V. 60, No. 1, (1981); Part 12, V. 60, No. 7, (1981), Part 13, V. 83, No. 11, (1983).
158. Tru-Tec Division of Koch Engineering *Co.* Inc. *Bulletin TPD-1 and TCS-1;* Koch Engineering Co. Inc, Wichita, Kn, 67208-0127.
159. Harrison, M. E., and J. J., Frana; "Trouble-Shooting Distillation Columns," *Chem. Eng.,* V. 96, No.3, p. 116, (1989).
160. Kister, H. Z., K. F., Larson and P. E., Madsen, "Vapor Cross-Flow Channeling on Sieve Trays: Fact or Myth?" *Chem. Eng. Prog.,* V. 88, No.11 p. 86, (1992).
161. Niesenfeld, A. E., "Reflux or Distillate: Which to Control?" *Chem. Eng.,* p. 169, Oct. 6, (1969).
162. Ludwig, E. E., *"Applied Project Engineering and Management,"* 2nd. Ed., Gulf Publishing Co., p. 433, (1988).
163. Rys, R. A., "Advanced Control Techniques for Distillation Columns," *Chem. Eng.,* p. 73, Dec. 10, (1984).
164. Kister, H. Z., *"Distillation Operation,"* McGraw-Hill Publishing Co., Inc., (1990).
165. Kister, H. Z., "How to Prepare and Test Columns Before Start-Up," *Chem. Eng.,* p. 97, Apr. 6, (1981).
166. Chin, T. G., "Guide to Distillation Pressure Control Methods," *Hydro. Proc.* V. 59, No. 10, p. 145, (1979).
167. Thurston, C. W., "Computer Aided Design of Distillation Column Controls," *Hydro. Prog.,* 60, No. 8, p 135, (1981).
168. Holland, C. D., *"Fundamentals of Multicomponent Distillation,"* McGraw-Hill Book Co. Inc., (1981).
169. Holland, C. D. and A. I., Liapis, *"Computer Methods for Solving Dynamic Separation Problems,"* McGraw-Hill, Inc., (1983).
170. Haring, H. G., B. J., Grootenhuis, and H. *W.,* Knol, "Programming Batch Distillation," *Chem. Eng.,* p. 159, Mar. 16, (1964).
171. Nisenfeld, A. E. and C. A., Stravinski, "Feed Forward Control for Azeotropic Distillation," *Chem. Eng.,* p. 227, Sept. 23, (1968).
172. Othmer, D. F., "Azeotropic Separation," *Chem. Eng. Prog.,* V. 59, No. 6, p 67, (1963).
173. Fair, J. R. and W. L., Bolles, "Modern Design of Distillation Columns," *Chem. Eng.*, p. 156, Apr. 22, (1968).
174. Kirkbride, C. G., *Petro. Refiner;* V. 23; No.9, p. 87, (1944).
175. Winn, F. W., "Equilibrium K's by Nomograph," *Petro. Refiner V*. 33, No. 6, p. 131, (1954).
176. Gerster, J. A., "A New Look at Distillation Tray Efficiencies," *Chem. Eng. Prog.,* V. 59, No.3, p. 35, (1963).
177. Williamson, W. R. and H. L. W. Pierson, senior thesis, University of Delaware, June, (1961).
178. Proctor, J. F., "A New Look at Distillation-2: Sieve and Bubble Plates Comparative Performance," *Chem. Eng. Prog,.* V. 59. No.3, p. 47, (1963).
179. Strand, C. P., "A New Look at Distillation-3: Bubble Cap Tray Efficiencies," *Chem. Eng. Prog.,* V. 59, No. 4, p. 58 (1963).
180. Sakata, M., "Liquid Mixing in Distillation Columns," *Chem. Eng. Prog.,* V. 62, No. 11 p. 98., (1966).
181. Hughmark, G. A., "Point Efficiencies for Tray Distillations," *Chem. Eng. Prog.,* V. 61, No.7, p. 97, (1965).
182. Bowman, J. D., "Use Column Scanning Predictive Maintenance," *Chem. Eng. Prog.,* V. 87, No.2, p. 25, (1991).
183. Fair, J. R., "How to Predict Sieve Tray Entrainment and Flooding," *Petro/Chem Engr.,* p. 45, Sept., (1961).
184. Kister, H. Z. and J. R., Haas, "Predict Entrainment Flooding on Sieve and Valve Trays," American Institute of Chemical Engineers, *Chem. Eng. Prog.,* V. 86, No. 9, p. 63, (1990).
185. Fair, J. R. and R. L. Matthews, *Petro. Ref.,* V. 37, No. 4, p. 153, (1958).
186. Kister, H. Z. and J. R., Haas, *I. Chem. Eng. Symp. Series*, V. 104, p. A 483, (1987).
187. Ward, T. J., "A New Correlation for Sieve Trays," *Chem. Eng.,* V. 96, No.6, p. 177, (1989).
188. Capps, R. W., "Consider the Ultimate Capacity of Fractionation Trays," *Chem. Eng. Prog.,* V. 89, No.3, p. 37, (1993).
189. Barnicki, S. D. and J. F., Davis, "Designing Sieve Tray Columns-Part 1," *Chem. Eng., V.* 96, No. 10, p. 140, (1989); and Part 2, *ibid No.* 11, p. 202, (1989).
190. Kister, H. Z., "Downcomer Design for Distillation Tray Columns," *Chem. Eng.,* p. 55, Dec. 29, (1980).
191. Yeoman, N., "The FRI Tray Data Sheet," *Chem. Eng. Prog.,* V. 85, No.12, p. 14, (1989).
192. Chase, J. D., "Sieve Tray Design, Part 1," *Chem. Eng.*, p. 105, July 31, (1967).
193. Fair, J. R., Chapter 15 in *Design of Equilibrium Stage Processes,"* B. D. Smith, McGraw-Hill, (1963).
194. Lessi, A., "How Weeping Affects Distillation," *Hydro. Proc.,* V. 51, No.3, p.109, (1972).
195. Jamison, R. H., "Internal Design Techniques," *Chem. Eng. Prog.,* V. 65, No. 3, p. 46, (1969).
196. Kister, H. Z., "Design and Layout for Sieve and Valve Trays," *Chem. Eng.,* V. 87, No.18, (1980).
197. Economopoulos, A. P., "Computer Design of Sieve Trays and Tray Columns," *Chem. Eng* V. 85, No. 27, p. 109, (1978).

198. Frank, J. C., G. R. Geyer, and H., Kehde, "Styrene-Ethyl-Benzene Separation with Sieve Trays," *Chem. Eng. Prog.*, V. 65, No. 2, p. 79, (1969).
199. King, C. J., *Separation Processes,* McGraw-Hill Book Co., p. 555, (1971).
200. Henley, E. J. and J. D., Seader, *"Equilibrium Stage Separations in Chemical Engineering,"* John Wiley and Sons, p. 507, (1981).
201. Klein, G. F., "Simplified Model Calculates Valve-Tray Pressure Drop," *Chem. Eng.,* V. 89, No. 9, p. 81, (1982).
202. Glitsch, Inc. "*Ballast Tray Design Manual,"* Bul. 4900, 5th. Ed. Glitsch, Inc., Dallas, Texas (1974, 1989).
203. Koch Engineering, Inc. *"Flexitray Design Manual,"* Bul. 960-1 (1982), Koch Engineering Co. Inc., Wichita, KN., (1982).
204. Nutter Engineering, " *Nutter float Valve Design Manual,"* and " *Electronic Design Manual Version* 2.0," Nutter Engineering, a Harsco Corp. Tulsa, Okla., (1990).
205. Bolles, W. L., *Chem. Eng. Prog.,* V. 72, No. 9, p. 43 (1976).
206. Thorngren, J. T., *Ind. Eng. Chem. Proc. Des.,* V. 11, No. 3, p. 428, (1972).
207. Thorngren, J. T., "Valve Tray Flooding Generalized;" *Hydro. Proc.,* V. 57, No. 8, p. 111, (1978).
208. Billet, R., "Development or Progress in the Design and Performance of Valve Trays," *British Chem. Eng.,* Apr, (1969).
209. Biddulph, M. *W.,* C. P., Thomas, and A. C., & Burton, "Don"t Downplay Downcomer Design," *Chem. Eng. Prog.,* V. 89, No. 12, p. 56, (1993).
210. Hsieh, C-Li., and K. J., McNulty, "Predict Weeping of Sieve and Valve Trays," *Chem. Eng. Prog.,* V. 89, No. 7, p. 71, (1993).
211. Fair, J. R., "How to Design Baffle Tray Columns," *Hydro. Proc.,* V. 72; No. 5, p. 75, (1993).
212. Lemieux, E. J., "Data for Tower Baffle Design," *Hydro. Proc.,* V. 62, No. 9, p. 106, (1983).
213. Kister, H. Z., K. F., Larsen, and P.E., Madsen, "Vapor Cross-flow Channeling on Sieve Trays, Fact or Myth?" *Chem. Eng. Prog.,* V. 88, No. 11, p. 86, (1992).
214. Winter, J. R., "Avoid Vibration Damage to Distillation Trays," *Chem. Eng. Prog.,* V.89, No. 5, p. 42, (1993).
215. Hasbrouck, J. F., J. G., Kunesh, and V. C., Smith, "Successfully Troubleshoot Distillation Towers," *Chem. Eng. Prog.,* p. 63, Mar., (1993).
216. Campagne, W. V. L., "Use Ponchon-Savarit in your Process Simulation, Part *1," Hydro. Proc.,* V. 72, No. 9, p. 41, (1993).
217. *Ibid,,V.72; No.* 10, p. 63, (1993).
218. Liebert, T., "Distillation Feed Preheat- Is It Energy Efficient?" *Hydro. Proc.,* V. 72, No. 10, p. 37, (1993).
219. U.S. Chemical Safety and Hazard Investigation Board- Final Investigation Report, Chevron Richmond Refinery Pipe Rupture and Fire, Report No. 2012-03-I-CA, January 2015.
220. Baker, James *et al.*, The Report of the BP U.S. Refineries Independent Safety Review Panel", 2007, http://www.safetyreviewpanel.com/, January 30, 2007.
221. American Society of Mechanical Engineers (ASME), "Rules for Construction of Pressure Vessels", 2001, ASME Boiler and Pressure Ves-sel Code, Section VIII, Division 1.
222. API, "Sizing, Selection, and Installation of Pressure-Relieving Devices in Refineries, Part I- Sizing and Selection", API Recommended Practice 520, 7th ed., 2000.
223. API, "Guide for Pressure -Relieving and Depressuring Systems", API Recommended Practice 521, 4th ed., 1997.
224. API, "Sizing, Selection and Installation of Pressure-Relieving Devices in Refineries, Part II- Installation", API Recommended Practice, 520, 4th ed., 1994.
225. Ashis Nag, Distillation & Hydrocarbon Processing Practices, PennWell, 2016.
226. N. Lieberman, Fundamentals of refinery safety – A study of process safety considerations arising from a raffinate splitter explosion, PTQ E3, pp 19 – 21, www.eptq.com, 2019.
227. Song, Liming,: Li, Baoan, Sirkar, Kamalesh, K., Jack L. Gilron, Direct Contact Membrane Distillation-Based Desalination: Novel Mem-branes, Devices, larger Scale Studies and Model, Ind. & Eng. Chem. Research, 46 (8): 2307-2323, 2007.
228. Srisurichan, S., Jiraratananon, R., A. G. Fane, Mass transfer mechanisms and transport resistances in direct contact membrane distillation process, Journ. of Membrane Science, 277 (1-2): 186-194, 2006.
229. Warsinger, D., Swaminathan, Jaichander, John Lienhard, Effect of Module Inclination Angle on Air Gap Membrane Distillation, Proceedings of the 15th International Heat Transfer Conference. 2014.
230. Warsinger, D., Swaminathan, Jaichander; Morales, Lucien, L, Lienhard, V., H John, Comprehensive condensation flow regimes in air-gap membrane distillation: Visualization and Energy Efficiency, Journ. of Membrane Science.
231. Karl Kolmetz, and T. M. Zygula, (Unpublished) Resolving Process Distillation Equipment Problems, Prepared for Presentation at the 5th Annual Regional Olefins Conference, Oct. 31st November, 3rd, 2000.
232. Stankiewicz., A, The Principles and Domains of Process Intensification, CEP, pp 23 – 28, March 2020.
233. Stankiewicz., A and J. A. Moulijn; Process Intensification: Transforming Chemical Engineering, Chem. Eng. Prog., 96 (1), pp. 22 – 34, Jan., 2000.
234. Ranshaw, C., Higee Distillation – An Example of Process Intensification, The Chem. Eng., Issue 389, pp 13 – 14, Feb. 1983.

235. Staak, D., *et al.*, Dividing Wall Column for Industrial Multipurpose Use, Chem. Eng. and Processing: Process Intensification, 75, pp 48-57, Jan. 2014.
236. Kiss, A. A., Maleta, V. N., Shevchenko, A., and O. Bedryk, Cyclic distillation: A novel enhanced technology for processing hydrocarbons and derivative, Hydro. Proc. Pp 33 – 38, January 2021.
237. Ozturk, G., Kibar, S., and G. Akyildiz, Troubleshoot flooding problems in a crude distillation column, Hydro. Proc., pp 57-60, Feb. 2021.
238. Sloley, A. W, and Christian Geipel., Questions & Answers, Distillation Roundtable, Engineering Practice, International Association of Certified Practicing Engineers, pp 5 – 24, Vol. 7, Number 25, March 2021.

Bibliography

American Petroleum Institute, Technical Data Book- Petroleum Refining, Port City Press Inc., Baltimore, Mid, (1966).

API data book procedure 11A1.1.

API data book procedure 11A1.4.

ASTM specification D-2161.

ASTM test method specification D976-80.

Ballard, Don, “How to Improve Glycol Dehydration”, Coastal Chemical Co., Inc., (1979).

Bandini, S.; Gostoli, C.; Sarti, G. C., Separation efficiency in vacuum membrane distillation, Journ. of Membrane Science, 73 (2-3): 217 – 229, 1992.

Berg, L., Chem. Eng. Prog., 65 (9), 52, (1969).

Blackwell, W. W., “Chemical Process Design On A Programmable Calculator”, McGraw-Hill Book Co., New York, (1984).

Boston, J. P., Britt, H. I., Jirapongphan, S., and V. B. Shah, “Foundations of Computer-Aided Chemical Process Design, 2, Ed., by R. S. Mah, and W. D. Seider, New York, American Institute of Chemical Engineers, 203, (1981).

Branan, C., “Rules of Thumb for Chemical Engineers”, 4th Ed., Gulf Publishing Professional, (2005).

Characterization of Petroleum Fractions, Ind. Eng. Chem., 27, 1460, (1935).

Chattopadhyay, P., “Unit Operations of Chemical Engineering Vol. II”, Khanna Publishers, (1998).

Chung, Hyung Won; Swaminathan, Jaichander, Warsinger, David, M; John H. V. Lienhard, Multistage vacuum membrane distillation (MSVMD) systems for high salinity applications, Journ. of Membrane Science. 497: 128 – 141, 2016.

Coker, A. K., “Equilibrium Flash Calculations Quickly Computed on PC,” Oil & Gas J., pp 55-58, January (1991).

Couper, J. R., W. R. Penny, J. R. Fair, and S. M. Walas, Chemical Process Equipment: Selection and Design, 2nd. Ed., Elsevier, (2005).

DeDoes, A. J., Goetz, M. D. and M. J. Misovich, “ A Simple Approach to Vapor Pressure Prediction”, CEP, pp 39-44, January (2007).

Dhole, V. R., and B. Linnhoff, “Distribution Column Targets,” Computers Chem. Eng., 17: 549, (1993).

Dimian, A. C., and C. S. Bildea, Chemical Process Design – Computer Aided Case Studies, Wiley –VCH, (2008).

Dodge, B. F., Chemical Engineering Thermodynamics, McGraw-Hill, New York, (1944).

Doherty, M. F and J. D. Perkins, Chem. Eng. Sci., 33, 281-301, (1978).

Doherty, M. F. and M. F. Malone, Conceptual Design of Distillation Systems, McGraw-Hill, New York, (2001).

Douglas, J. M. and A. Jafarey, “Short-Cut Techniques for Distillation Column Design and Control. 2, Column Operability and Control, Ind. Eng., Chem. Process Des. Dev., vol. 18, No. 2, pp 203-210, (1979).

Douglas, J. M., The Conceptual Design of Chemical Processes, McGraw-Hill, New York, (1988).

Edmister, W. C., “Applied Hydrocarbons Thermodynamics”, Gulf Publishing Co., Houston, Texas, (1961).

Elliott, J. R. and Carl T. Lira, “Introductory Chemical Engineering Thermodynamics”, Prentice Hall Int. Series., (1998).

Engineering Data Book, Gas Processing Suppliers Association, Vol. 2, 12th ed., Section 19, Tulsa, Oklahoma (2004).

Ewell, R. H., J. M. Harrison, Lloyd Berg, Ind. Eng. Chem., 36 871, (1944).

Fenske, M. R., “Fractionating of Straight-Run Pennsylvania Gasoline”, Industrial Engineering Chemistry, Vol. 24, No. 5, pp 482, (1932).

Flower, J. R., and M. A. Jackson, “Energy Requirements in the Separation of Mixture by Distillation,” Trans. Scheme., 42: T249, (1964).

Francis, S., Manning and E. Richard, Thompson’s Oilfield Processing of Petroleum, Volume One: Natural Gas. PennWell Books, (1991).

Gallant, R. W., *Hydrocarbon Processing*, V. 47, No. 8, p. 127, (1968).

Haddad, H. N., and D. B., Manley, "Improve crude oil fractionation by distributed distillation", Hydrocarbon Processing, pp. 93-108, May (2008).

Hadden, S. T., and H. G. Grayson, Petrol Refiner, p. 207, September (1961).

Hines, A. L. and R. N. Maddox, Mass Transfer Fundamentals and Applications, Prentice-Hall, Inc., New Jersey, (1985).

Hirata, M. Ohe Shuzo and Kunio Nagahama, "Computer Aided Data Book of Vapor-Liquid Equilibria, Kodansha Ltd. Elsevier Scientific Publishing Co.

Horsley, L. H., "Azeotropic Data III", in Advances in Chemistry Series, American Chemical Society, Washington, D. C., Vol, 116, (1973).

Humphrey, J. L. and A. F. Siebert, "New Horizons in Distillation", Chem. Eng., pp. 88-98, Dec., (1992).

Hydrocarbon Processing: Refining Processes 2006; Gas Processing, 2006; Petrochemical Processing, 2005, www.hydrocarbon-processing.com.

Jafarey, A., Douglas, J. M., and T. J. Mcvay, "Short-Cut Techniques for Distillation Column Design and Control. 1, Column Design", Ind. Eng., Chem. Process Des. Dev., vol. 18, No. 2, pp 197-202, (1979).

Kaes, G. L., Refinery Process Modeling: "Guidelines Can Help Choose Proper Process For Gas Treating Plants", Oil & Gas Journ., Janu-ary 10, (1977).

Karanikola, Vasiliki, Corral, Andrea, F, Jiang, Hua; Eduardo A. Saez, Ela, Wendell, P., Arnold, Robert G., Sweeping gas membrane distil-lation: Numerical simulation of mass and heat transfer in a hollow fiber membrane module, Journ. of Membrane Science. 483: 15-24, 2015.

Kayihan, F., "Optimum Distribution of Heat Load in Distillation Columns Using Intermediate Condensers and Reboilers," AIChE Symp. Ser., 192 (76):1, (1980).

Khayet, M; Cojocaru, C; A. Baroudi, Modeling and Optimization of sweeping gas membrane distillation. Desalination. 287: 159-166.

Khoury, F. M., Multistage Separation Processes, 3rd. ed., CRC Press.

Kister, H. Z., Component Trapping in Distillation Towers: Causes, Symptoms and Cures, CEP, p. 22, August 2004.

Kister, H. Z., Distillation Troubleshooting, AIChE®- Wiley Interscience, 2006

Kister, Henry, Z., "Distillation Design", McGraw-Hill, (1992).

Knapp, J. P., and M. F. Doherty, Ind. Eng. Chem. Res., 31, 346-357, (1992).

Kostecke, S. T., "Speed of Hand Calculator Programs Can Be Improved," Oil & Gas J., p. 107, August, (1980).

Ludwig. E. E., "Applied Process Design for Chemical and Petrochemical Plants", Volume 2, 3rd. Ed., Gulf Publishing Company, Hou-ston, TX, (1997).

Luyben, W., Distillation Design and Control Using Aspen Simulation, John Wiley & Sons, (2006).

Maddox, R. N., "Gas and Liquid Sweetening Campbell Petroleum Series, (1997).

Manjoorana, Koch. B. S., "Modern Petroleum Chemistry: An overview", Kannatheri Publications, Karla, India, (2003).

Matar, S. and L. F. Hatch, Chemistry of Petrochemical Processes, 2nd. Ed., Gulf Publishing Co., Houston, TX, (2001).

Maxwell, J.B., Data book on Hydrocarbons, Princeton, N.J., Van Nostrand, (1968).

McWilliams, M. L., "An Equation to Relate K-factors to Pressure and Temperature", Chem. Engineering, 80 (25), 138, Oct. 29, (1973).

Meili, A., Heat pumps for distillation, Chem. Eng., Prog., No. 6, vol. 86, p. 60-65, June (1990).

Meszaros, I and D. Hangi, Sulzer Technical Review, Nr. 1, (1999).

Meszaros, I., "Environment-friendly and economical distillation with heat pump", Summaries 2 – Separation Process, 18th Int. Cong. Of Chemical and Process Eng., p. 3.13, CHISA (2008).

Meszaros, I., and A. Meili, Heat integration and heat pump technology in relation to distillation, 6th IEA Heat Pump Conference, Berlin (1999).

Nelson, W. L., "Regulation of I.B.P. and E. P.", Oil & Gas Journ., p 143, August 25, (1945).

Nelson, W. L., Oil & Gas Journ., pp 125, March 25, (1968).

Nutter Float Valve Design Manual, Nutter Engineering Co., Tulsa, Oklahoma, August (1981).

Oliver, I. H., "Hand Calculator Program Speeds Flash Calculations," Oil & Gas J., p. 130, March 31, (1980).

Parkash, S., "Refining Processes Handbook", Gulf Professional Publishing: Elsevier, (2003).

Pena, M. Diaz, Colin, A. C., and A. Compostizo, J., Chem. Thermodyn, vol. 10, pp 337-341, (1978).

Perry, J. H. (editor), *Chemical Engineers' Handbook*, McGraw- Hill Book Co., New York, (1963).

Perry, R. H., and Green Don, W and J. O. Maloney, "Perry's Chemical Engineers' Handbook", 7th ed., Mc. Graw-Hill Book Co., (1999).

Pham, H. N. and M. F. Doherty, "Design and Synthesis of Heterogeneous Azeotropic Distillation- III. Column Sequences", Chem. Eng. Sci., 45, (1990).

Pitzer, K. S., D. Z. Lippman, R. F. Curl, Jr., C. M. Huggins, and D. E. Petersen, J., Am. Chem. Soc., 77, 3433-3340, (1955).

Poling, B. E., J. M. Prausnitz, and J. P. O'Connell, "The Properties of Gases and Liquids", 5th Ed., McGraw-Hill, New York, (2001).

Reid, R. C., Prausnitz, J. M. and B. E. Poling, "The Properties of Gases and Liquids", 3rd. Ed., McGraw-Hill Book Co., NY., (1977)., Reid, *et al*. 4th Ed., Mc Graw-Hill, New York, (1987).

Ryan, J. M., "Repotting The McCabe-Thiele Diagram", Chemical Engineering, pp 109-113, May, (2001).

Seader, J. D and Ernest, J. Henley, "Separation Process Principles", 2nd Ed., John Wiley & Sons Inc., (2006).

Seader, J. D., and E. J. Henley, Separation Process Principles, 2nd Ed., John Wiley & Sons, Inc. (2006).

Shah, K. K. and G. Thodos, Ind. Eng. Chem., 57 (3), 30-37, (1965).

Shinskey, F. C. "Distillation Control", McGraw-Hill, p 43, New York, (1977).

Shreve, R. N. and G. T. Hatch, Chemical Process Industries, 5th Ed., McGraw-Hill, New York, (1984).

Smith, J.M., van Ness, H. C. and M. M. Abbott, "Introduction to Chemical Engineering Thermodynamics", 7th Ed., McGraw-Hill Int., (2005).

Smith, R., and Linnhoff, B., "The Design of Separators in the Context of Overall Processes", Trans. IChemE, ChERD, 66: 195, (1988).

Smith, R., Chemical Process Design and Integration, John Wiley & Sons Ltd, (2007).

Smoker, E. H., "Analytic Determination of Plates in Fractionating Columns", Trans. Ache, vol. 34, p 165, (1938).

Sorel, M., "La Rectification de l'alcool', Paris, (1893).

Stichlmair, J. G. and J. R. Fair, Distillation: Principles and Practice, Wiley-VCH, (1998).

Stichlmair, J., J. R. Fair, and J. L. Bravo, Chem. Eng. Progr., 85 (1), 63-69, (1989).

Swaminathan, Jaichander; Chung, Hyung Won; Warsinger, David, M; John H. V. Lienhard, Energy efficiency of permeate gap and novel conductive gap membrane distillation, Journ. of Membrane Science. 502: 171- 178, 2016.

Tassios, D. (ed.), "Extractive and Azeotropic Distillation", Am. Chem. Soc. Adv. Chem. Ser. No., 115, (1972).

Tennyson, R. N. and R. P. Schaap, "Guidelines Can Help Choose Proper Process For Gas Treating Plants", Oil & Gas Journ., January 10, (1977).

Thakore, S. B., and B. I. Bhatt, Introduction to Process Engineering and Design, Tata Mc Graw-Hill, New Delhi, (2007).

Van Dongen, D. B., and M. F. Doherty, "On the Dynamics of Distillation Processes", Chem. Eng. Sci., 39, (1984).

Walas, S. M., "Phase Equilibria In Chemical Engineering", Butterworths Publishers, Boston, (1985).

Wankat, P. C., "Separation Process Engineering", 2nd Ed., Pearson Education, Inc., (2007).

Widago, S and W. B. Seider, "Azeotropic Distillation", AIChE Journal, Vol. 42, No. 1, pp. 96-130, (1996).

Wilke, C. R., Chang, P., *AIChEJ*., V. 1, p. 264, (1955).

Wunder, J. W., "How to Design a Natural Gas Drier", Oil & Gas Journ., August 6, pp 137-148, (1962).

20

Packed Towers and Liquid–Liquid Extraction

Packed Towers

Packed towers are used increasingly in a variety of applications in the chemical process industries. The reasons for this are an increase in the availability of design information, the evolution of higher-capacity and higher-efficiency mass transfer packings, and improvements in distributors and support plates. Today, packings can be considered for most services where high numbers of theoretical stages are required for mass transfer.

Most of the fractionating columns in gas-processing plants that were previously equipped with trays can now be filled with packings. Packings such as slotted rings in metal and plastic and plastic saddles offer large improvements in capacity for absorption and heat transfer applications. An example for this is the use of slotted rings in a gas-plant demethanizer, which has resulted in a significant capacity improvement over trays.

Packed towers are used increasingly in small fractionating towers and absorbers where access is restricted. Their use is also advantageous where corrosion control is essential. Because of liquid-distribution concerns, packing still is usually not recommended for large-diameter columns or where a high turndown ratio is required. All packings tend to spread liquid as if it flows downwards. This characteristic is adequate in dispersing the liquid between distribution points but not to correct poor distribution across a tower without considerable loss of efficiency. Studies of liquid spread in tower packing have been reported by Hoek [1]. Figure 20.1A shows a cross section of a typical packed tower; Figure 20.1B illustrates a typical packed tower with internals for improved distillation. Figures 20.1C and 20.1D show a photograph of a packed column and a schematic of the pilot plant packed column, respectively. The shell is usually cylindrical, although square wooden, light metal, or reinforced plastic towers are used. The basic unit consists of

1. shell
2. packing (one or more sections)
3. packing support(s)
4. liquid distributor(s)
5. intermediate supports and redistributors
6. gas and liquid entrance and exit nozzles

Many of the mechanical aspects of tower construction and assembly have an influence on the design and interpretation of tower performance. Every effort should be made to increase the effectiveness of contact between the process streams and to reduce losses by entrainment or wall effects while minimizing pressure drop. At the same time, the design must be consistent with the economics dictated by the process and type of construction.

A. Kayode Coker. Petroleum Refining Design and Applications Handbook Volume 3, (703–960) © 2022 Scrivener Publishing LLC

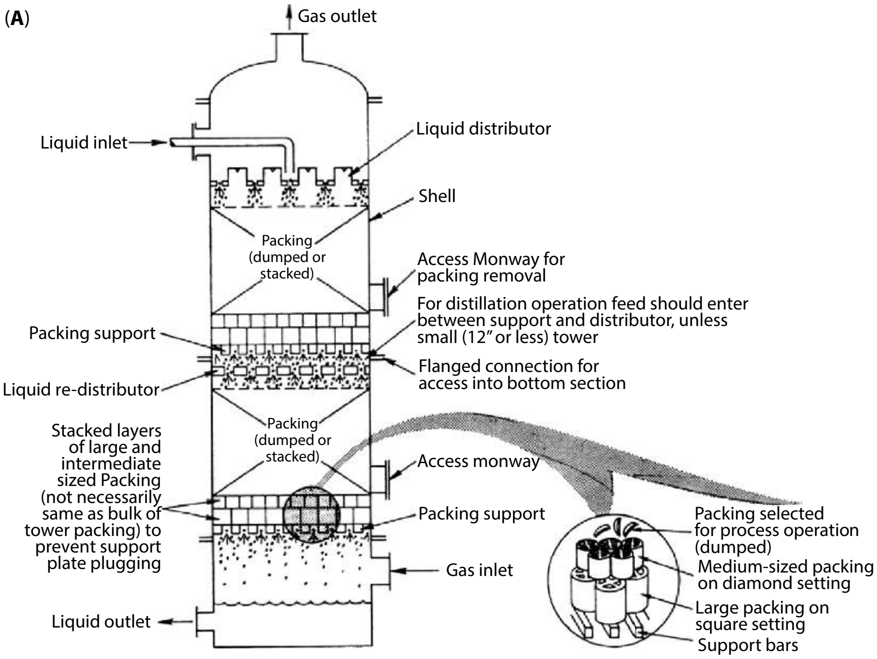


Figure 20.1A Cross-section of typical packed tower.

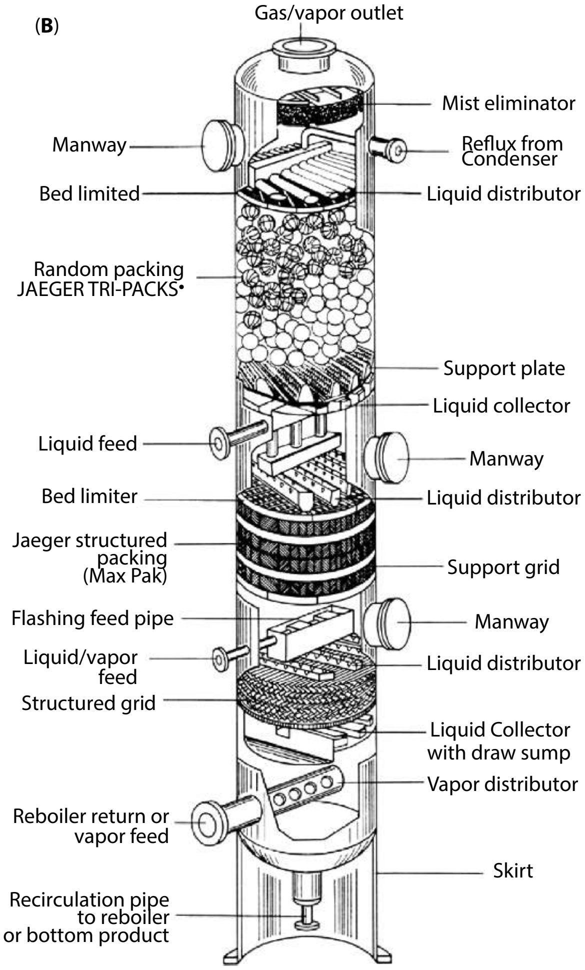


Figure 20.1B Typical packed tower with internals for improved distillation. Used by permission of Jaeger Products Inc. Bull. 1100.

Figure 20.1C Photograph of a pilot plant absorption column (courtesy of Armfield, U.K.).

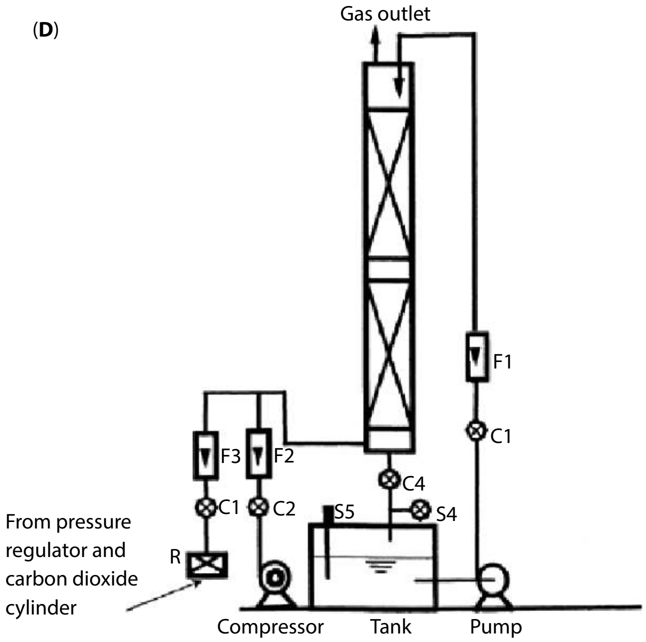


Figure 20.1D Schematics of a pilot plant absorption column (courtesy of Armfield, U.K).

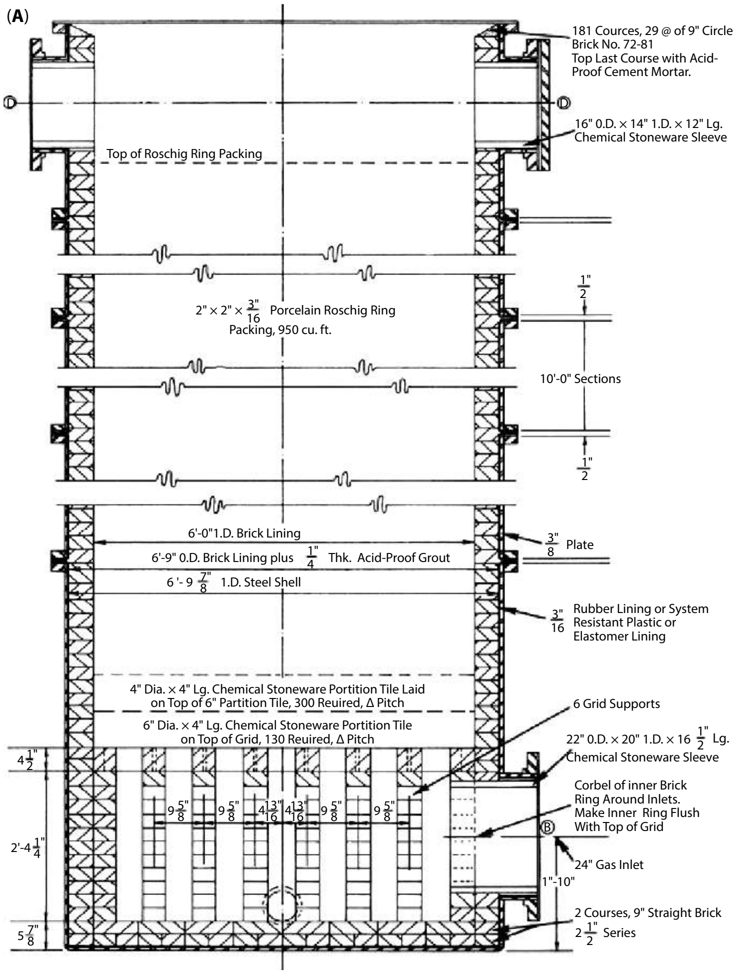


Figure 20.2A Cross section of membrane and brick-lined packed tower. Depending on tower diameter, certain dimensions should be modified, particularly parking. Also see Figures 20.2B and 20.2C.

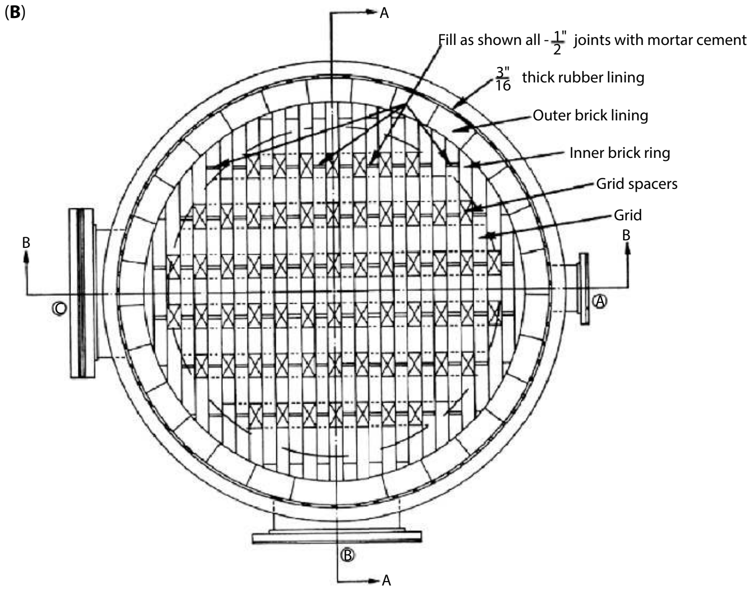


Figure 20.2B Cross-section of membrane and brick-lined packed tower. Depending on tower diameter, certain dimensions should be modified, particularly parking.

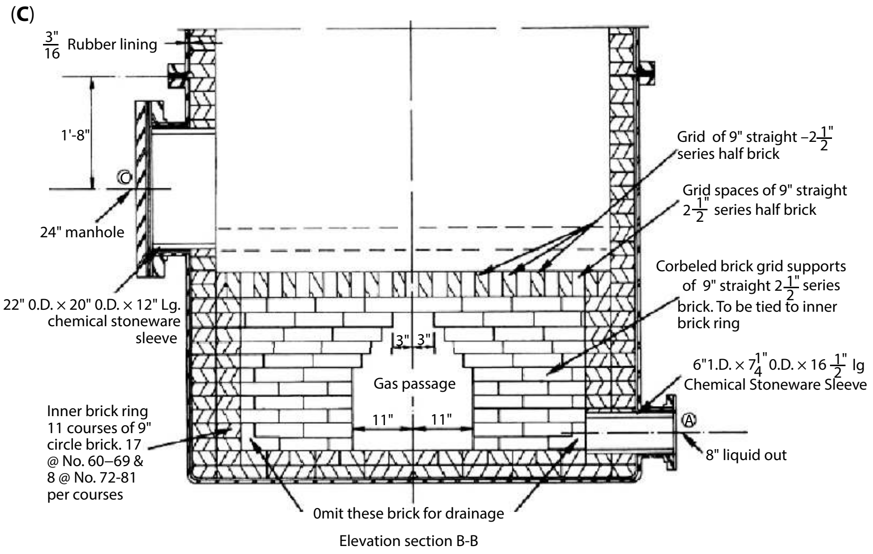


Figure 20.2C Cross-section of membrane and brick-lined packed tower. Depending on tower diameter, certain dimensions should be modified, particularly parking.

Packing in packed columns is not designed by the process designers engaged in overall unit design. It is designed by packing experts based on the input of vapor–liquid flows and physical properties such as liquid–vapor densities, viscosities, and so on. However, the process engineer is required to determine the diameter and height of the column, ensuring that it must be adequate for column internal suppliers to install the internals without any problem. Both the height and diameter of the column are estimated before the column internals are designed by column internal experts. Diameter can be estimated by a steady-state simulator (e.g., Hysys, UniSim, Pro II, etc.) that takes trays or packing into account, or more conservatively, diameter can be determined only by taking the trays into account. The height is determined by height equivalent to a theoretical plate (HETP). The author has developed a computer program to determine these parameters as illustrated in this book.

20.1 Shell

The shell may be of metal (steel, alloy, or non-ferrous), plastic, wood, or some combination, which may require the addition of liners or inner layers of rubber, plastic, or brick. The mechanical problems of attaching inner nozzles, supports, and brick require considerable attention that is not an integral part of sizing the equipment. Figures 20.2A–C show a typical large steel brick-lined-membrane lined tower with corbeled brick support locations. In these towers, temperature and/or corrosive conditions usually dictate the composition of internal lining, and the requirements for the proper acid- (or alkali) proof cements.

Ceramic, plastic, and other non-metal tower shells are used quite often (Figures 20.3, 20.4, and 20.5). It is important in ceramic construction to consider that the main inlet or outlet nozzles or any other large connections should be oriented 90° to each other in order to reduce the possibility of cracking the walls, as most cracks go one-half diameter. Preferably there should only be one nozzle at any one horizontal plane. The nozzles should never carry any piping or other stress load.

The bell and spigot type tower, Figures 20.3 and 20.4, is satisfactory for 2 to 2.5 psi in 12-in. dia. to 30-in. dia. towers when the joints are packed with chevron-type caulking compound. For an operating pressure of 5 psi in 18-in. through 48-in. dia. tower, use of non-asbestos and silica cement. Special hold-down packing gland-type rings will allow this pressure to be raised slightly. Porcelain towers should be used for the higher pressures rather than the weaker stoneware.

The rate of heating or cooling a stoneware or porcelain tower should not exceed 15°F/min.

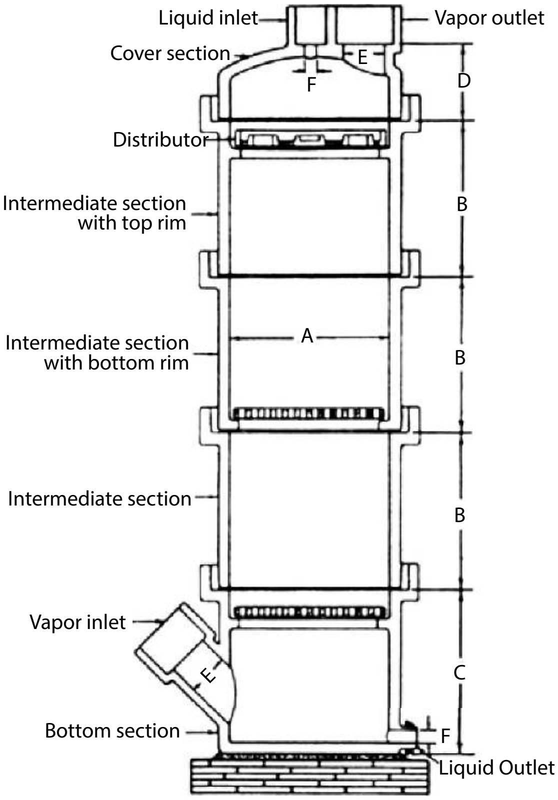


Table of dimensions (inches)

	INSIDE DIAMETER				
A	8	12	15	18	24
B	370. 30"4 34" UAI-60"				
C	15	19	24	30	30
D	6	7	7	7	7
E	3	4	4	6	6
F	1	1	1½	2	2

Figure 20.3 Bell and spigot ceramic tower. Used by permission of General Ceramics and Steatite Corp.

20.2 Random Packing

The distributor and packing define the performance of this equipment. Their proper selection entails an understanding of packing operational characteristics and the effect on performance of the points of significant physical difference between the various types.

Good progress has been made in the past decade in the development of packing types for difficult and wide-ranging process applications. These types include:

1. ***Random particle*** packings are discrete, individually shaped particles designed to provide contacting surfaces between (normally) down-flowing liquid and up-flowing vapor/gas. The degree of effectiveness of the various shapes varies along with the mass pressure drop through the packed bed. Usually these particles are "dumped" into the column (tower) and allowed to gently float to their free-fall resting position in a column full of water. Some shapes and sizes are not installed using water, but dumped in using a special "sock" that allows the particles to be lowered without a damaging free fall. Sometimes large particles are hand set "dry" into position to tightly fill the tower. See Figure 20.1A. The vapor–liquid performance is different for each method of loading, and appropriate data must be available to properly size the tower. The random "dumped" (wet or dry) method is the technique usually used for most published data (see Figures 20.6A–20.6X).
2. ***Structured packing***, which is offered by several manufacturers, is usually composed of pack "pads" fabricated by shaping/crimping, bending, rolling, etc. sheets of thin gauge metal or wire (see Figures 20.6Y to 20.6OO). Some "pads" or packs are formed using various plastics material, selected to be resistant to the fluid services involved.
3. ***Grid*** packing is probably the newest packing type. It is lower in pressure drop and has higher capacity and lower efficiency than the other types (see Figures 20.6PP to 20.6UU).

DIMENSIONS OF STANDARD TOWERS

Inside Diameter of Tower A	24 in.	30 in.	36 in.	40 in.	48 in.	60 in.
A	24	30	36	40	48	60
A^1	23¼	29¼	35	39	47½	58
B	27½	34	41½	45	54½	67½
C	36	36	36	36	36	36
C^1	30	30	36	36	40	40
D	6	6	6	6	6	6
E	15	15	15	15	15	15
F	10	12	14	14	18	18
G	6	8	12	12	18	18
H	8	10½	14¾	14¾	21	21
I	1	1	1	1½	1½	1½
J	2½	2½	2½	3	3	3
L	4	4	4	4	4	4
M	3	3	3	3	3	3
O	7	9⅓	11½	11 5/3	11¾	18
P	3	3	3	3	3	3
Q	7	8	11	11	14	16
R	1¼	1¼	1½	1½	1¾	2
R^1	1½	1 5/3	1¾	1¾	2	2
S	¾	¾	1	1	1¼	1¼

S is inside diameter of ground in faucet.

WEIGHTS

	Diameter					
	24 in.	30 in.	36 in.	40 in.	48 in.	60 in.
Tower Cover, Fig. 360 Weight, lbs.	80	130	215	300	500	700
Tower Cover Distributor, Fig. 401 and 401a Weight, lbs.	86	60	100	120	170	250
Plain Tower section, Fig. 356 Weight, lbs.	220	300	430	520	700	1000
Tower Section, with Rim Fig. 356a and 356b Weight, lbs.	225	310	445	540	750	1100
Supporting Plate, Fig. 279 Weight, lbs.	30	80	140	160	250	400
Tower Section, with branch, Fig. 355 Weight, lbs.	230	320	530	630	900	1800
Tower Saucer, Fig. 354 Weight, lbs.	120	200	250	340	500	800
Tower Bottom Section, Fig. 355a Weight, lbs.	260	370	610	730	1000	1500

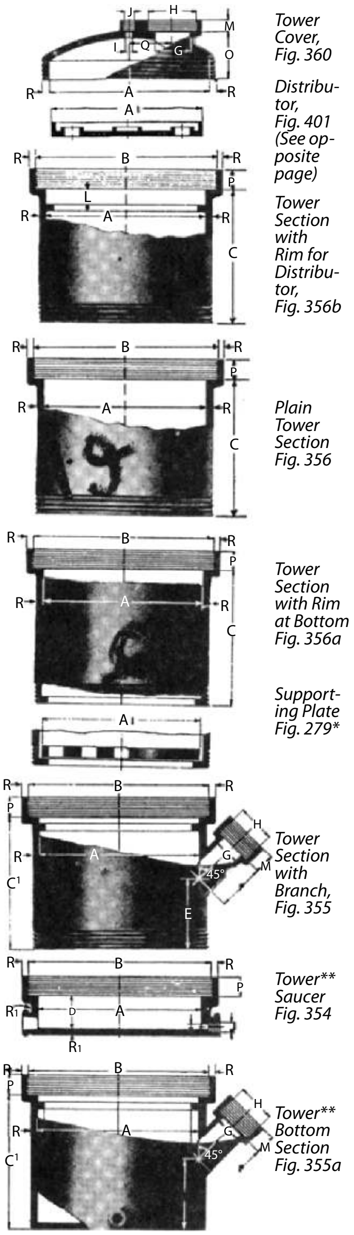


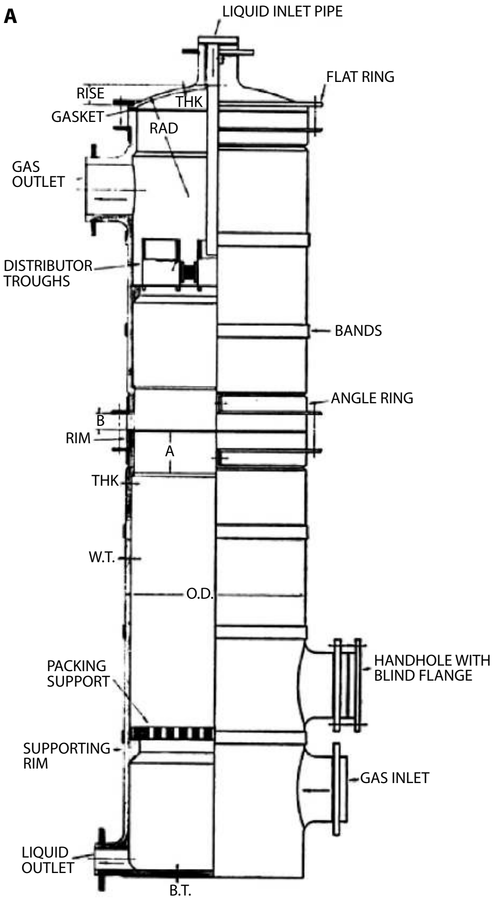
Figure 20.4 Physical dimensions of stoneware tower sections, bell and spigot. Used by permission of General Ceramics and Steatite Corp.

The types and corresponding physical data for packing are given in Figures 20.6 and Tables 20.1 through 20.15. Evaluation of these materials in various conditions of service is provided later. However, Table 20.16 outlines packing service applications, and Table 20.17 summarizes usual applications of each packing type.

20.3 Packing Supports

The packing support may be anything from cross-grid bars spaced to prevent fall-through of packing to more refined specialty units designed to direct the flow of gas and liquid (see Figures 20.7A–20.7F). Good tower performance is definitely linked to proper packing support. The net free flow cross-sectional area of the support should be 65% (or larger) of the tower area and greater than the free area of the packing itself. In addition, the effect of the free area “blocking” by the positioning of the packing on the support must be considered. To allow for this, every effort should be made to obtain as large a support-free area as possible and yet remain consistent with the structural strength of the material

A



B

DESIGN INFORMATION FOR SINGLE OR MULTIPLE SECTION CYLINDRICAL HAVEG TOWERS

Size	O.D.	Max. Height of Section	Wall & Bottom Thickness		Support Rim	Joint for Section Towers										COVER							Staves and Hoops	
						Rim			Angle Ring			Gasket		Bolts		Head			Flat Ring			Bolts		
			W.T.	B.T.	I.D. × Tk.	A	B	Th'k.	Size	I.D.	B.C.	I.D.	O.D.	No.	Size	Rad.	Rise	Th'k.	I.D.	O.D.	Th'k.	Size	Size Hoops	No. of Staves
12"	13½	15'-0"	3/8	¾	……	4	1	1 3/8	2 ×2 ×3/18	12 7/8	15 1/8	11¼	13¼	12	½×3½	8¾	2¼	½	11¼	16¾	¾	½×3		
15"	16½	15'-0"	3/8	¾	11 3/8×1	4	1	1 1/8	2 ×2 ×3/18	15 3/8	17 7/8	14	16¼	12	½×3½	12½	2½	½	14	19½	¾	½×3		
18"	19½	15'-0"	3/8	¾	14 ×1	4½	1½	1 2/8	2 ×2 ×3/18	18 3/8	20 7/8	17	19¼	12	½×4½	16½	2	½	17½	23	½	½×3½	……	……
20"	21½	15'-0"	3/8	¾	16¼×1¼	4½	1½	1 1/8	2 ×2 ×¼	20 7/4	23 1/8	19¼	21½	16	½×4½	18	2¾	½	19¼	24¾	½	½×3½	……	……
2'	25¼	15'-0"	3/8	¾	20 ×1¼	4½	1½	1¼	2 ×2 ×¼	24½	26¾	23	25¼	16	½×4½	25	2 7/8	3/8	23½	28½	½	½×3½	……	……
2'-6"	31½	15'-0"	¾	1	24 ×1½	5	1½	1¼	2½×2½×7/16	30¾	33½	29	31½	20	½×5	32 3/8	3¼	3/8	29	36	5/8	½×4	……	……
3'	37¼	14'-6"	¾	1	29¾×1½	5	1½	1¼	2½×2½×7/16	36½	39¼	34¾	37¼	28	½×5	39	3 7/8	¾	34¾	41¾	5/8	½×4	……	……
3'-6"	43½	14'-6"	¾	1	36 ×1½	6½	3	1¼	2½×2½×7/16	42¾	45½	41½	43½	28	½×8	47¾	4 3/8	¾	41	48	5/8	½×5½	5/8	25
4'	49½	14'-6"	1	1¼	40¾×2	7	3	1 3/8	3 ×3 ×½	48½	51½	46¼	49½	32	5/8×8	48½	6	¾	47	54	¾	5/8×6	5/8	28
5'	61½	14'-0"	1	1¼	53½×2	7	3	1 5/8	3 ×3 ×½	60½	63½	58¼	61½	40	5/8×8	58 5/8	7¼	1	59	65	¾	5/8×6½	5/8	35
6'	74	14'-0"	1	1¼	65 ×2½	7½	3	1 5/8	3½×3½×½	73	76¼	70¾	74	48	5/8×8	75	8¾	1	71½	79½	¾	5/8×6½	5/8	42
7'	86 3/8	13'-6"	1¼	1 3/8	……	7½	3	2	3½×3½×½	85 3/8	88 3/8	82 3/8	86¾	52	5/8×8	87½	10	1	83 3/8	91 3/8	7/8	5/8×6½	¾	49
8'	98 3/8	13'-0"	1¼	1 3/8	……	8	3	2	4 ×4 ×3/8	97 3/8	100 7/8	98 3/8	98 3/8	60	¾×8½	91¼	12¼	1¼	95 3/8	104 3/8	7/8	¾×7	¾	56
9'	110 3/8	12'-6"	1¼	1 3/8	……	9	3	2	5 ×5 ×3/8	109 1/8	112 7/8	106 3/8	110 3/8	72	¾×8½	108 1/8	13 1/8	1¼	107 3/8	117 3/8	7/8	¾×7	¾	63
10'	120	12'-0"	1½	1 3/8	……	9	3	2¼	5 ×5 ×¾	118¾	122½	115½	120	80	¾×8½	119½	15	1¼	116 3/8	126 3/8	7/8	¾×7	¾	69

All dimensions are in inches unless otherwise noted.

Figure 20.5 Typical reinforced plastic tower construction; with dimension used by permission of Haveg Corp., Bull. F-7.

being used. If this area is too restricted, liquid build-up will occur at the plate, reducing efficiency and increasing pressure drop of the tower, which could lead to a flooding condition. A lot depends on the material of construction that the system requires; for example, carbon or graphite bar grids, brick grid piers, some steel grating grids and most rubber or plastic covered metal grids have inherently, low free cross-sectional areas. These may be less than 65% free area. Pressure drops through support plates, such as shown in Figure 20.7D, are reported [2, 82] not to exceed 0.3 in. water for most applications. Also see Figures 20.7A–C, E, and F. Fouling service can create pressure drop problems.

In some large towers, the support grid is built up from supporting brick arches coming from the bottom (see Figures 20.2A, B, and C). Quite often in large towers, drip point grid tile is used as the supporting first layer, either as a support "plate" itself or as the support for other packing stacked on it (Table 20.15). This initial stacking of the first and perhaps second courses of packing prevents the blocking of free area usually associated with dumping packing

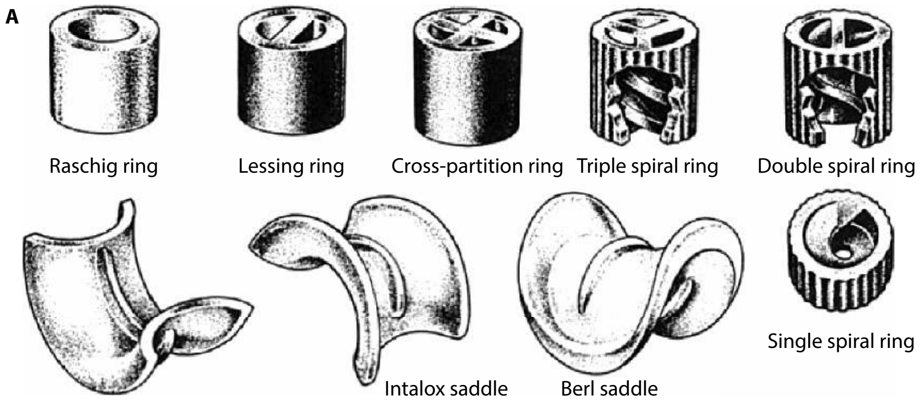


Figure 20.6A Various packing shapes (ceramic).

Figure 20.6B Raschig rings (ceramic, carbon, metal).

Figure 20.6C Intalox® saddles (ceramic), dumped. Used by permission of Norton Chemical Process Products Corp.

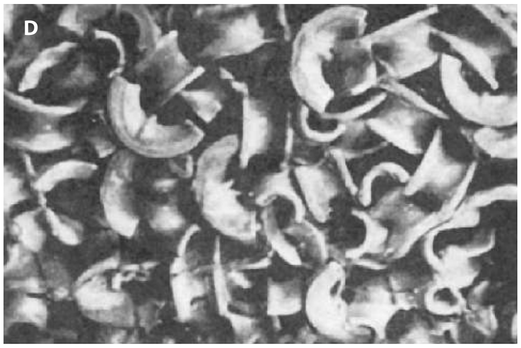


Figure 20.6D Berl saddles (ceramic), dumped.

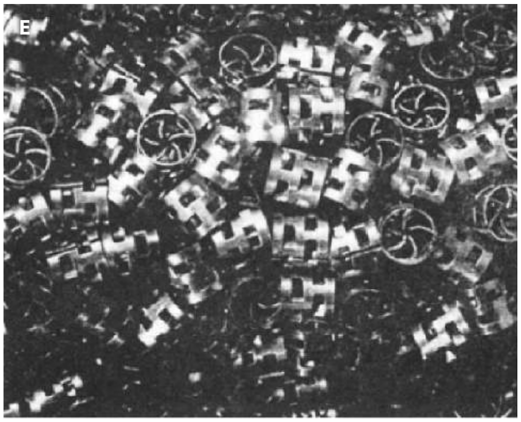


Figure 20.6E Pall rings (metal), dumped.

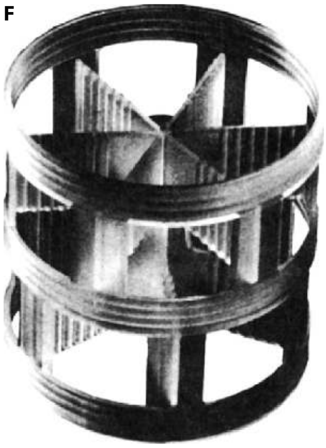


Figure 20.6F Koch Flexring (plastic). Used by permission of Koch Engineering Co., Inc., Bull. PFR-1.

Figure 20.6G Norton Hy-Pak® (metal). Used by permission of Norton Chemical Process Products Co., Ind. Bull. DC-11 and MTP-1 (4/94).

H

Figure 20.6H Plastic pall ring. Note: Glitsch Ballast Ring® and others are quite similar. Used by permission of Norton Chemical Process Products Co. Inc. Bull. DC-11 and PTP-1 (11/87).

I

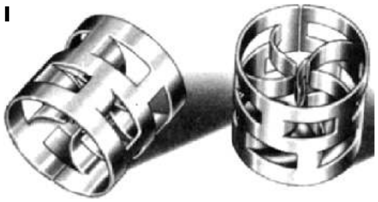
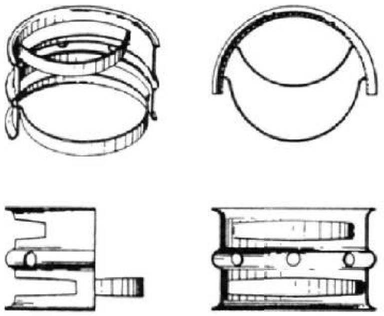


Figure 20.6I Metal pall ring. Note: Glitsch Ballast Ring® and Koch Engineering Flexring® are quite similar. Used by permission of Norton Chemical Process Products Co. Inc. Bull. N-60D and Bull. MTP-1 (4/94).

J

Figure 20.6J Metal Intalox (IMTP®). Used by permission of Norton Chemical Process Products Co. Inc. Bull. 1HP-1 (12/91).

K



MECHANICAL SPECIFICATIONS						
	No. 0.7	No. 10	No. 1.5	No. 2.0	No. 2.5	No. 3.0
Pieces/ft^3	4740	1900	760	385	250	120
ft^2/ft^3	69	51	38	29	25	20
lb/ft^3	11.0	11.1	11.3	10.8	9.0	8.3
% Void	97.8	97.8	97.8	97.9	98.2	98.4
Relative HETP Values	0.64	0.78	0.92	1.00	1.17	1.40

Figure 20.6K Nutter Ring™ (metal random packing). Used by permission of Nutter Engineering. Harsco Corp., Bull. NR-2.

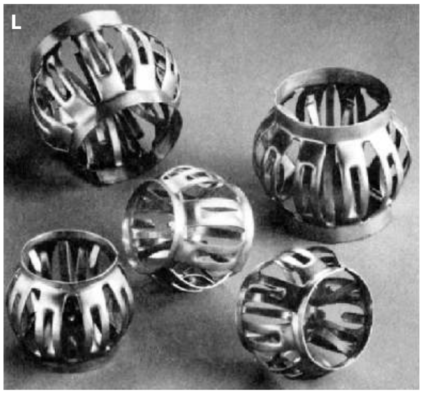


Figure 20.6L Metal Top-Pak® low liquid/high gas flow and high heat transfer packing. Used by permission of Vereinigte FullKorper-Fabriken GmBH & Co. Ransbach Baumbach, Germany.

M

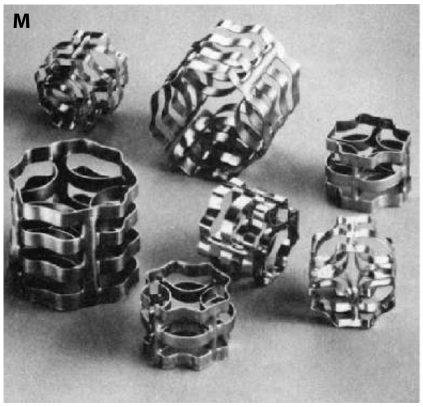


Figure 20.6M Metal VSP® high capacity packing. Used by permission of Vereinigte FullKorper-Fabriken GmBH & Co. Ransbach Baumbach, Germany.

N(a)

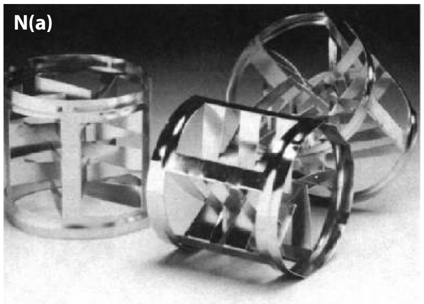


Figure 20.6N(a) Koch Metal HcKp™ random packing. Used by permission of Koch Engineering Co., Inc. Bull. KRP-2.

N(b)

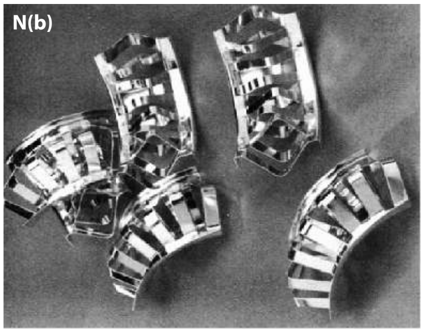


Figure 20.6N(b) Fleximan™ high – performance random metal packing; available in two sizes and reported by manufacturer to be a fourth-generation random packing. Used by permission of Koch Engineering Co., Inc., Bull. KFM-1.

O

Figure 20.6O Koch Metal Flexring®; available in 5/8-in. to 3½.-in. diameter. Used by permission of Koch Engineering Co., Inc. Bull. KRP-2.

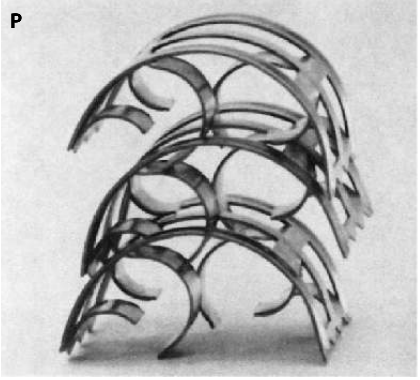


Figure 20.6P Metal Chempak® packing. Used by permission of Chem-Pro Equipment Corp., licensed from Dr. Max Leva.

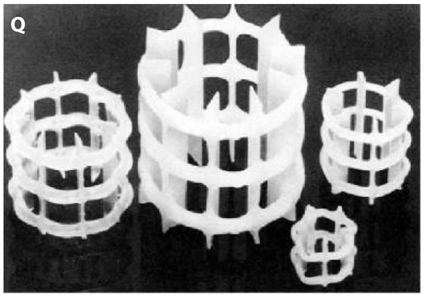


Figure 20.6Q Plastic Nor-Pac® packing; fabricated of most corrosion-resistant plastics. Used by permission: NSW Corporation.

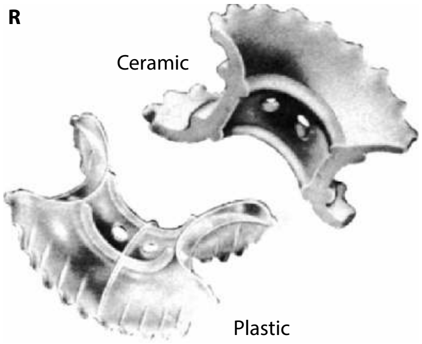


Figure 20.6R Super Intalox® saddles. Note: Glitsch Ballast saddle® are quite smaller. Used by permission of Norton Chemical Process Products Corp., Bull. N-6OD.

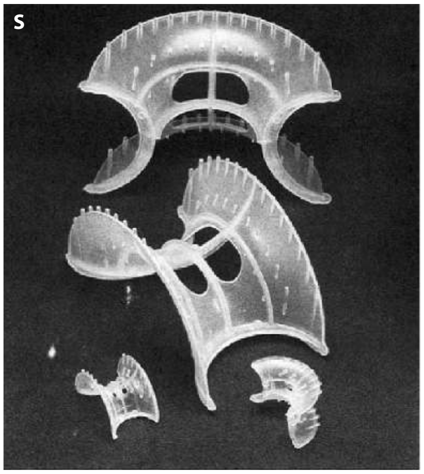


Figure 20.6S Koch Flexisaddles™ (plastic); available in 1-in. to 3-in. Used by permission of Koch Engineering Co., Inc., Bull KRP-2.

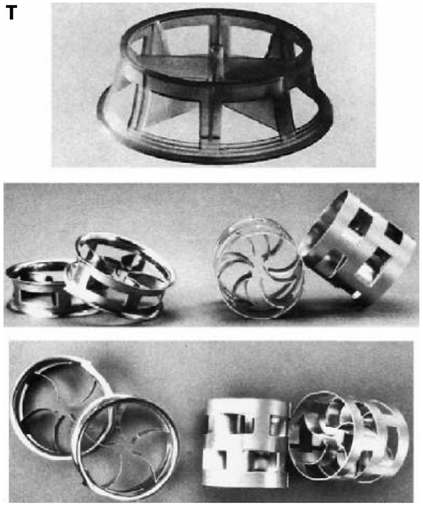


Figure 20.6T (Top) Cascade® Mini-Ring, (metal and plastic). Originally used by permission of Mass Transfer, Inc., now Glitsch Inc. (middle and bottom) Elevation and plan views of Ballast® rings (right) and Cascade Mini-Rings (left). Note how high aspect ratio of former permits occlusion of interior surfaces. Low aspect ratio of Cascade Mini-Rings, on the other hand favors orientation that exposes internal surfaces for excellent film formation, intimate mixing, and gas-liquid contact. Used by permission of Glitsch Inc., Bull. 345.

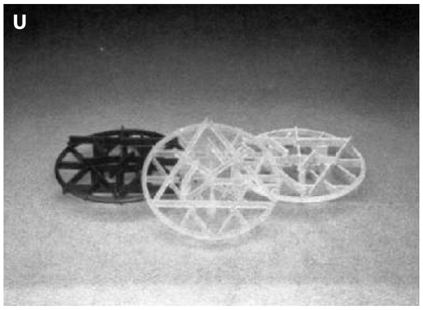


Figure 20.6U Norton Intalox® high-performance Snowflake® packing (plastic). Used by permission of Norton Chemical Process Products Corp., Bull. 1SPP-1R.

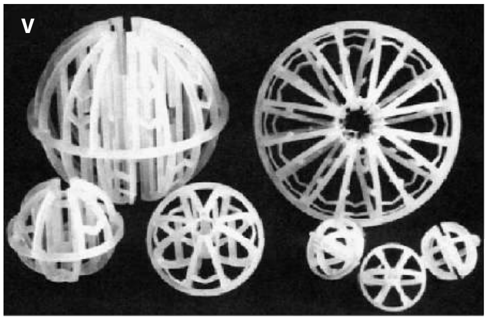


Figure 20.6V Jaeger Tri-Packs® high-performance fabricated of corrosion-resistant plastics. Used by permission of Jaeger Products, Inc.

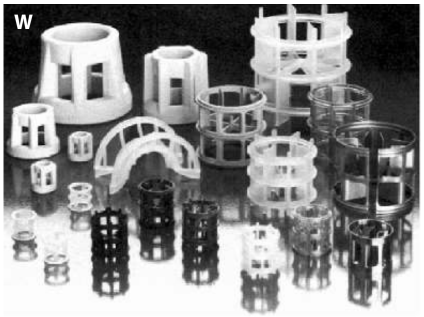


Figure 20.6W Rauschert Hiflow® high – performance rings and saddles packing (plastic and metal). Used by permission of Rasuchert Industries, Inc., div. of Rauschert GmbH & Co., KG; Paul-Rauschert-Str., 6; D-96349 Steinwiesen, Germany.

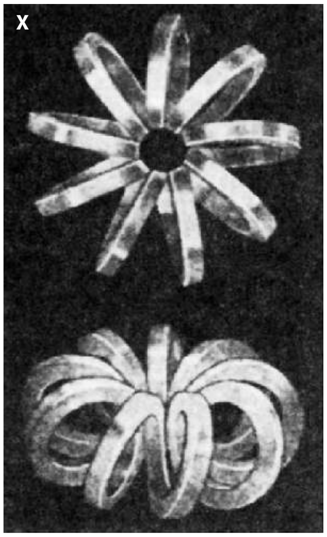


Figure 20.6X Teller rosette packing (plastic). Used by permission of Hawshaw Chemical Co., and Dr. A. J. Teller.

Figure 20.6Y Panapak formed/structured packing for a 6-ft. diameter column (original design). Used by permission of Packed Column Corp.

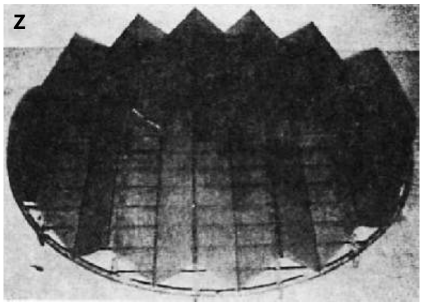


Figure 20.6Z Panapak packing for a 6- ft. diameter column, latest design. Used by permission of Packed Column Corp.

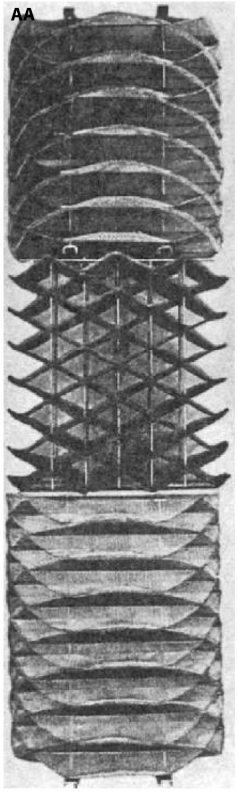


Figure 20.6AA Cartridge 8 ½ - ft. high of 32 in. I.D. Spraypak Packing. Used by permission of Denholme, Inc. Bull. No. 8.

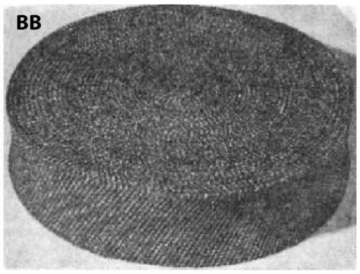


Figure 20.6BB Goodloe packing. Used by permission originally of Packed Column Corp., now under license to Glitsch Inc.

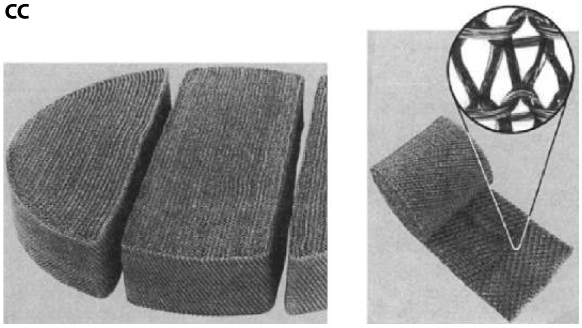
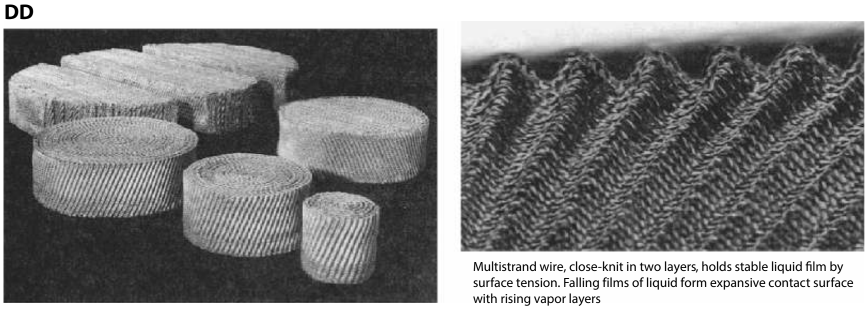


Figure 20.6CC York – Twist^{TM} structured tower packing. Note insert showing weave and sections fabricated to fit through manway for larger towers. Used by permission of Otto H. York Co., Inc., a division of Glitsch, Inc.



Multistrand wire, close-knit in two layers, holds stable liquid film by surface tension. Falling films of liquid form expansive contact surface with rising vapor layers

Figure 20.6DD ACS knitted mesh structured packing. Used by permission of ACS Industries, Inc., Separations Technology Group, Bull. B-129 (1992).

Figure 20.6EE Koch/Sulzer® packing. Used by permission of Koch Engineering Co. Inc., Bull. KS-2.

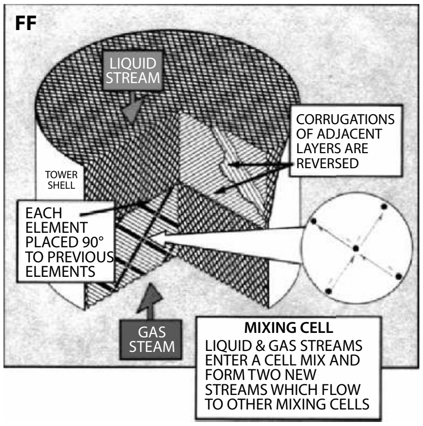


Figure 20.6FF Liquid and gas mixing in Koch/Sulzer® packing elements. Used by permission of Koch Engineering Co. Inc., Bull. KS-2.

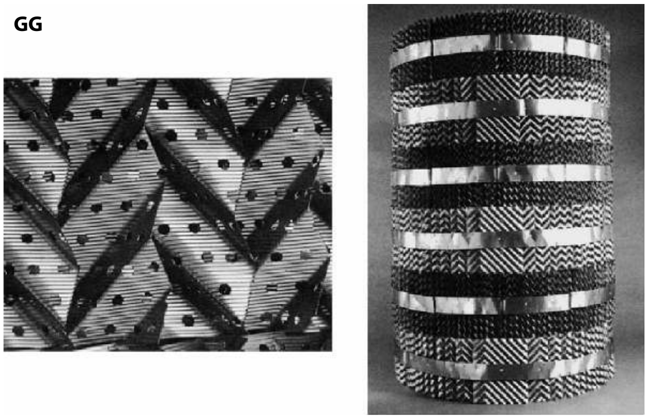


Figure 20.6GG Koch Flexipac® structured packing. Used by permission of Koch Engineering Co. Inc., Bull. KFP-4

Figure 20.6HH Metal Max-Pak™ structured packing fabricated from corrugated sheets. Used by permission of Jaeger Products, Inc. Bull. 500.

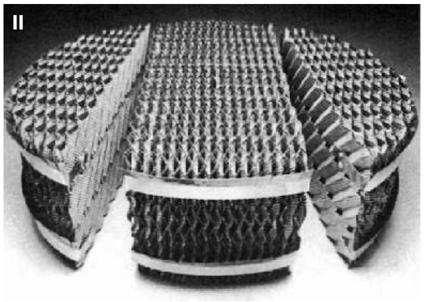


Figure 20.6II Intalox® high-performance structured packing. Used by permission of Norton-Chemical Process Products Corp., Bull. IS-2T.

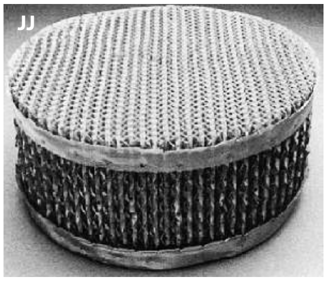


Figure 20.6JJ Intalox® high-performance wire gauze structured packing. Used by permission of Norton Chemical Process Products Corp., Bull. WG-1.

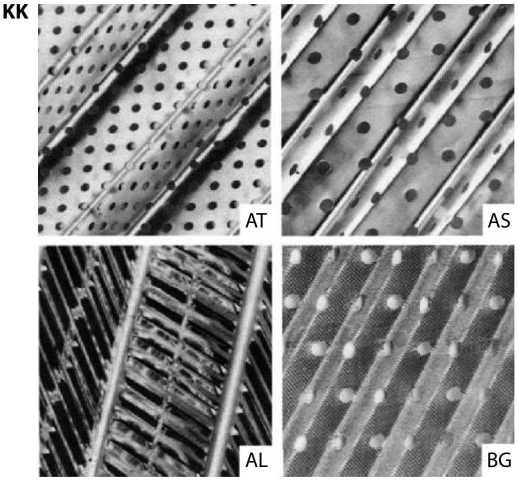


Figure 20.6KK Glitsch Gempak® structured packing series designs. Series AT-Smooth surface with small crossover points on narrow rectangular pitch. A general-purpose style, especially suitable for low wetting rates and high vacuum applications. Series AS-A general purpose style suited for high wetting and heat transfer applications. Both AT and AS series are particularly suited for refinery applications and can be used in the wash zones of atmospheric and vacuum crude towers, or FCC main fractionators. Series AL-Metal surface with louvers in a repetitive pattern. Especially suited for low wetting rates and chemical applications. Series BG-A wire gauze packing, crimped to 60° from the horizontal. Recommended for high efficiency applications at very low wetting rates (clean service). Used by permission of Glitsch, Inc. Bull. 5140 (1993).

LL

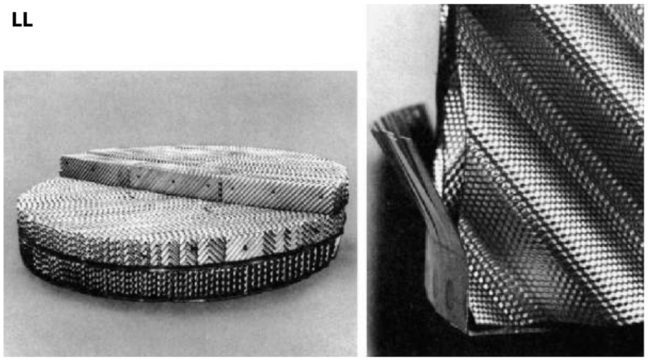


Figure 20.6LL Montz-Nutter high-efficiency structured packing assembly (Type B-1™ with wiper bands and seal strip). Used by permission of Nutter Engineering, Harsco Corp., Bull. B-1.

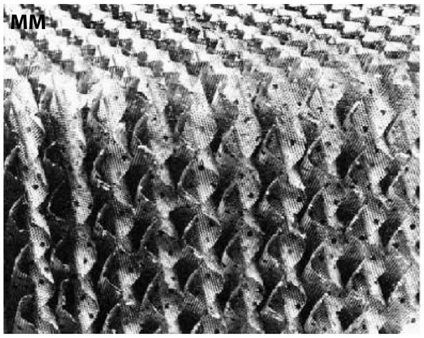


Figure 20.6MM Nutter BHS™ structured packing with expanded metal texture, which maintains high mass transfer efficiency across all operating conditions. Nutter Engineering designs and manufactures BSH in North America under exclusive license from Julius Montz GmbH of Germany. Used by permission of Nutter Engineering, Harsco Corp.

NN

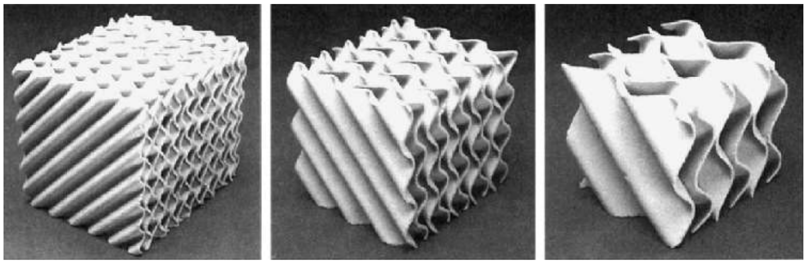


Figure 20.6NN Koch Ceramic Flexeramic® structured packing of stoneware or ceramic fabrication. Used by permission of Koch Engineering Co. Inc., Bull KCP-2.

OO

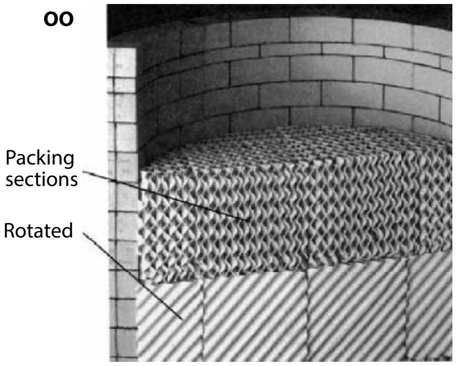


Figure 20.6OO Installation of Ceramic Flexeramic® structured packing of stoneware or ceramic. Used by permission of Koch Engineering Co. Inc., Bull. KCP-2.

PP

Figure 20.6PP Nutter Snap-Grid™ grid packing as assembled. Used by permission of Nutter Engineering, Harsco Corp. Bull. CSG-1.

QQ

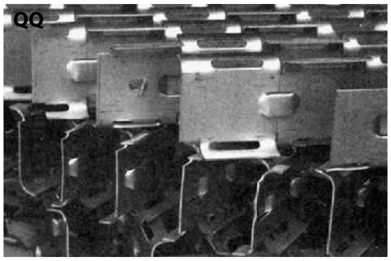


Figure 20.6QQ Nutter Snap-Grid™ high-capacity grid packing with good structural integrity (interlocking) and reduced fouling. Used by permission of Nutter Engineering, Harsco Corp. Bull. CSG-2.

RR

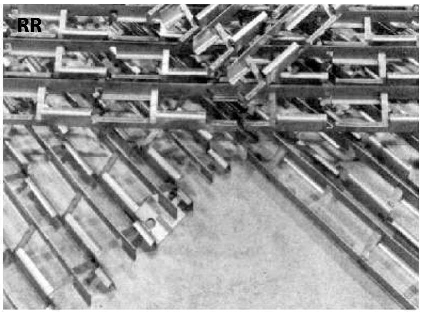


Figure 20.6RR Intalox® Grid-type packing useful for fouling, solids and heat transfer; high-strength, low-pressure drop, high capacity, better efficiency than most structured and random packing. Used by permission of Norton Chemical Process Products Corp., Bull. IG-1.

SS

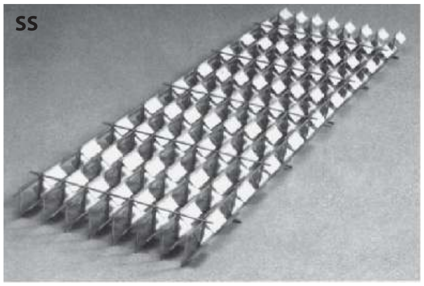


Figure 20.6SS Flexigrid® Style 3 high-efficiency structured packing. Used by permission of Koch Engineering Co. Inc., Bull. KFG-2.

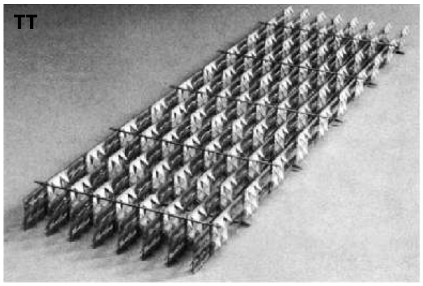
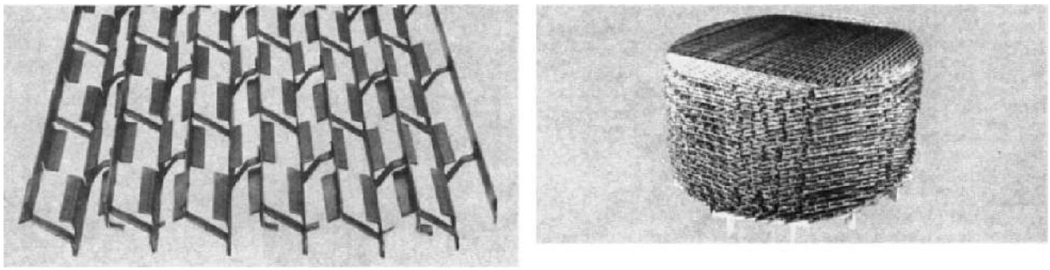


Figure 20.6TT Flexigrid® Style 2 high-efficiency structured packing. Used by permission of Koch Engineering Co., Inc., Bull. KFG-2.

UU



EF-25A Grid is used to process coking-prone feeds and up to 40% solids slurries

Because Grid is free to expand and contract without damage under thermal stress. Grid can maintain its structural integrity in turbulent vapor streams as hot as 1000°F

Figure 20.6UU Glitsch-grid™ packing, Style EF-25A. Used by permission of Glitsch, Inc., Bull. 163L84B.

Table 20.1 Chemical stoneware and chemical porcelain raschig rings.

Normal size, in.	Wall thickness in.	O.D and length, in.	Approx. avg. number per ft^3	Approx. avg. weight per ft^3, lb**	Approx. avg. surface area f^2/ft^3	Percent free gas space	Equivalent sphere. diam. dp, in.
½	3/32	½	10,700	52	124	65	0.48
5/8	3/32	5/8	5,600	4852	100	68	0.57
3/2	3/32	¾	3,100	46	78	69	0.65
1	1/8	1	1,350	42	58	72	0.87
1 ¼	3/16	1 ¼	670	46	44	71	1.10
1 ½	¼	1 ½	390	40	36	73	1.40
2	¼	2	165	39	28	74	1.75
3	3/8	3	50	35	19	77	2.65
+3(D)	3/8	3	74	67	29	60	2.65
+3(S)	3/8	3	64	58	25	66	2.65
4	½	4	25	40			

5/8, 2 ½, 4, and 6 in. Available on special order. The 3-in., 4-in., and 6-in. O.D. sizes are also made with ribbed or corrugated outside surfaces. The 3-in. and 6-in. sizes can be made in lengths up to 12 in., on special order.

**Porcelain Rings are about 5% heavier than Stoneware. These weights are the average for both.

+ Data for stacked arrangement. "D" indicated diamond pattern. "S" indicated square pattern.

Original by permission, M. Leva, Ref. 40 and Bulletin TP-54, U.S. Stoneware Co., Akron, OH.; updated CTP-1 dated 9/88, by permission of Norton Chemical Process Products Corp.

Table 20.2 Carbon raschig rings.

Nominal size, in.	Wall thickness, in.	O.D. and length, in.	Approx. avg weight per ft^3	Approx. avg. weight per ft^3, lb	Approx. avg. surface area f^2/ft^3	Percent free gas space	Equivalent sphere. diam. Dp, in.
¼	1/16	¼	85,000	46	212	55	0.27
½	1/16	½	10,600	27	114	74	0.42
¾	1/8	¾	3,150	34	74	67	0.72
1	1/8	1	1,330	27	57	74	0.87
1 ¼	3/16	1 ¼	678	31	45	69	1.10
1 ½	¼	1 ½	390	34	37.5	67	1.40
2	¼	2	166–157*	27	28.5	74	1.75
3	5/16	3	50–44*	23	19	78	2.50
3(D)	5/16	3	74	49.5	29	66	2.50
3(S)	5/16	3	64	43	25	71	2.50

S = Square pattern } Stacked
D = Diamond pattern }

From: M. Leva, U.S. Stoneware Co., Akron, Ohio, Ref. 40, and Bulletin CP-2512, National Carbon Co., New York, N.Y.; updated 12/93 by permission Norton Chemical Process Products Corp., Bull CTP-1.

*Varies with supplier.

Table 20.3 Metal raschig rings.

Nominal size, in.	Wall thickness, in.	O.D and length, in.	Approx. avg. number per ft^3	Approx. avg. weight per ft^3, lb*	Approx. avg surface area f^2/ft^3	Percent free gas space	Equivalent sphere. diam. Dp, in.
¼	1/32	¼	88,000	133	236	72	0.22
3/8**	1/32	3/8	27,000	94	–	81	–
½**	1/32	½	11,400	75	123	85	0.34
½**	1/16	½	11,200	132	118	73	0.44
5/8	–	5/8	5,800	–	–	86	–
19/32	1/32	19/32	7,300	66	112	86	0.40
19/32	1/16	19/32	7,000	120	106.5	75	0.50
¾**	1/32	¾	3,340	52	81.7	89	0.46
¾**	1/16	¾	3,140	94	70.6	80	0.58
1**	1/32	1	1,430	39	62.2	92	0.56
1**	1/16	1	1,310	71	55.2	86	0.70

(*Continued*)

Table 20.3 Metal raschig rings. (*Continued*)

Nominal size, in.	Wall thickness, in.	O.D and length, in.	Approx. avg. number per ft^3	Approx. avg. weight per ft^3, lb*	Approx. avg surface area f^2/ft^3	Percent free gas space	Equivalent sphere. diam. Dp, in.
1 ¼**	1/16	1 ¼	730	62	49.3	87	0.75
1 ½	1/16	1 ½	375	49	39.2	90	0.90
2**	1/16	2	165	37	29.3	92	1.15
3**	1/16	3	55	25	19.8	95	–
3(D)	1/16	3	74	35	29	93	–
3(S)	1/16	3	64	30	25	94	–

Usually metal Raschig rings are made with fitted butted-joints.
Original by permission, M. Leva, [40] and U.S. Stoneware Co., Akron, OH.; updated 9/88 by permission of Norton Chemical Process Products Corp.,
Bull. MTP-1 and TP-78.
Note: Sizes availability varies with manufactures.
S = Square pattern } Stacked
D = Diamond pattern } Stacked
*Based upon Carbon Steel Rings: other weight are: Stainless Steel 105%; Copper 120%; Aluminum 37%; Monel and Nickel 115%
**Usual commercial sizes.

Table 20.4 Metal lessing rings.

Size, in.	Wall thickness, in.	O.D and length, in.	Approximate number per ft^3	Approximate weight per ft^3, lb*	Approximate surface area f^2/ft^3	Percent free gas space	Equivalent spherical dia., dp, in.
¼	1/32	¼	81,840	195	306.8	60	0.24
3/8	1/32	3/8	25,110	114	217.1	76	0.31
½	1/32	½	10,974	100	166.4	81	0.38
½	1/16	½	10,230	172	153.4	66	0.46
5/8	1/32	5/8	6,789	86	145.6	82	0.45
5/8	1/16	5/8	6,510	150	138.4	68	0.57
¾	1/32	¾	3,171	71	108.5	85	0.52
¾	1/16	¾	2,967	130	93.3	71	0.63
1	1/32	1	1,339	52	81.5	90	0.62
1	1/16	1	1,251	95	73.7	80	0.77
1 ¼	1/16	1 ¼	674	81	64.1	82	0.99
1 ¾	1/16	1 ½	391	65	53.6	87	1.02

(*Continued*)

Table 20.4 Metal lessing rings. (*Continued*)

Size, in.	Wall thickness, in.	O.D and length, in.	Approximate number per ft^3	Approximate weight per ft^3, lb*	Approximate surface area f^2/ft^3	Percent free gas space	Equivalent spherical dia., dp, in.
1 ¾	1/16	1 ¾	246	58	46.0	89	1.13
2	1/16	2	167	49	40.8	90	1.24

All figures are on a dumped basis. Metal Lessing Rings are also made in stainless steel, copper, and aluminum.
*Weight shown is for carbon steel. By permission: U.S. Stoneware Co. Bul. TP54, now, Norton Chemical Process Product Corp., Stow, OH. [5] except column 8, [40].

Table 20.5 Ceramic lessing rings.

O.D and length, in.	Wall thickness, in.	Approx. no. of rings per ft^3, dumped	Approx. wt. per ft^3 dumped	Percent free gas space	Approx. surface area per ft^3	Equi. sphere. diameter dp, in.
1	1/8	1,300	50	66	69	0.95
1 ¼	3/16	650	56	62	53	1.20
1 ½	¼	350	58	60	40	1.55
2	3/8	150	49	68	32	1.90

From M. Leva, U.S. Stoneware Co., [40], by permission, now, Norton Chemical Process Products Corp., Stow, OH.

Table 20.6A Chemical stoneware and ceramic berl saddle packings.

Nominal Size	Approx. average number/ft^3	Approx. wt./ft^3, lb	Approx. average surface area ft^2/ft^3	Percent free gas space*	Equivalent spher. diam. dp, in.
¼	11,300	56	274	60–67	0.23
½	16,200	54	142	63	0.42
¾	5,000	48	82	66	0.58
1	2,200	45	76	69–70	0.76
1 ½	580	38	44	73–75	1.10
2	250	40	32	72–75	1.55

From M. Leva, U.S. Stoneware Co., [40], by permission.
*Updated by permission Jaeger Products, Inc.

Table 20.6B Steel* berl saddles.

1	2,500	87	85	83
1 ½	825	60	58	88

Other metals available.

Courtesy Maurice A. Knight Co., Bulletin No. 11, now, by permission Koch Engineering Co., Inc.

Table 20.7 Ceramic Intalox®* Saddles.

Nominal size, in.	Approximate number per ft^3	Approximate weight per ft^3, lb	Approximate surface area ft^2/ft^3	Percent free gas space	Equivalent spherical diameter dp, in.
¼	117,500	54	300	64	0.20
3/8	50,000	50	–	67	–
½	17,000	46	190	69	0.32
¾	6,200	44	102	71	0.48
1	2,100	42	78	72	0.68
1 ½	660	39	59.5	74	0.96
2	240	37	36	75	1.38
3	52	37	–	75	–

Data shown applies to Intalox Saddle made either from chemical stoneware or chemical porcelain. Weights per cubic foot are based on chemical porcelain. Chemical stoneware Intalox Saddles will weigh approximately 5% less.

*Trade name Norton Co.

Orig. By permission: U.S. Stoneware Co., Akron, Ohio, Ref. 5, except column 6, from Ref. 40; updated 9/88 by permission of Norton Chemical Process Products Corp., Inc., Bull. CTP-1, 9/88.

Table 20.8 Ceramic cross-partition rings.

Outside diameter in.	Outside length in.	Wall thickness, in.	Number of ringers per ft^3		Weight of ringers per ft^3, lb		Percent free gas space		Ft^2 of surface area per ft^3 of packing		Net cross-section area of packing in ft^2/ft^2	
			D setting	S setting	D setting	S setting	D setting	S setting	D setting	S setting	D setting	S setting
3*	3	3/8	74	50*–64	73	47*–63	47–48	54–55	41–43	35–37	.53	.46
4	3	½	41	36	81–72	71–63	46–49	52–55	31–33	27–29	.54	.47
4	4	7/16	31	27	81–62	71–54	45–56	52–61	30–32	26–29	.54	.47
6	4	5/8	14	12	73–70	62–60	51–50	58–57	22	19	.49	.42
6	6	5/8	9	8	70	62	53–50	58–56	20	18	.49	.42

Also made with outer surfaces ribbed or corrugated, and in lengths up to 12 in. Rings with different wall thickness than above can be made on special order. Porcelain rings weigh about 5% more than above. For D and S patterns, see Figure 9.16. Compiled from Ref. 40, U.S Stoneware Bull. TP-54, and Maurice A. Knight Co. (now, Koch Engineering Co., Inc.) Chemical Equipment Bulletin, by permission; updated 2/89 by permission of Norton Chemical Process Products Corp. Bull. WG-1.

For 3-in. size only.

*Indicates dumped rings only; all others stacked diamond (D) or square pattern stacked (S).

Table 20.9 Ceramic spiral packing rings*.

	Outside diam in.	Outside length in.	Wall thickness, in.	Number of ring per ft^3		Weight of ring per ft^3		Percent free gas space		Ft^2 of surface area per ft^3 of packing		Net cross-section area of packing in ft^2/ft^2	
				D setting	**S setting**	**D setting**	**S setting**	**D setting**	**S setting**	**D setting**	**S setting**	**D setting**	**S setting**
Single	31/4–3	3	5/16	63–74	54–64	60–67	52–58	58–52	66–59	40–41	34–36	.32	.27
Spiral	6	6	5/8	31	27	61–60	55–52	60–57	67–64	32	28	.33	.28
	3 ¼	3	5/16	63	54	67	58	56	63	44	37	.37	.32
Double	4	4	5/8	31	27	64	58	59	64	35	31	.38	.33
Spiral	6	6	½	9	8	65	58	64	68	23	21	.32	.29
	3 ¼	3	5/16	63	54	69	60	50	57	50	42	.51	.44
Triple	4	4	5/8	31	27	65	59	53	58	40	35	.46	.40
Spiral	6	6	½	9	8	68	60	60	64	24	21	.32	.29

*Basic data in table for U.S Stoneware “Cyclohelix” spiral packing, Bul. TP 54, Ref. 5. Data for other spiral packings shown set to right from Maurice A. Knight Co. (now, Koch Engineering Co., Inc.) Bulletin No. 11, by permission. For D and S patterns, Figure 14.6.

Table 20.10 Pall rings—metal.

Size O.D & length, in.	Wall† thickness, in.	Approx. number ft³	Approx.* weight, lb/ft³	Surface area ft²/ft³	Percent free gas space
5/8	0.018	5,800	37	104	93
1	0.024	1,400	30	63	94
1 ½	0.030	375	24	39	95
2	0.036	165	22	31	96
3 ½	0.048	33	17	20	97
Plastic‡					
5/8	–	6,060	5.95		87
1	–	1,430	4.4		90
1 ½	–	420	4.35		91
2	–	175	3.85		92
3 ½	–	33	3.45		92

By permission. Norton Chemical Process Products Corp., TP-78 and PR-16 and MTP-1 (9/88): other manufacturer's data are equivalent

†Standard gauge carbon steel, approximate.

‡Weights referenced to polypropylene: other plastic available, high-density polyethylene, glass reinforced polypropylene and fluorinated vinyls.

Table 20.11 Teller Rosette (Tellerette) plastic*.

Nominal size, in.	No. Units per ft³	Weight per ft³, lb	Surface area ft²/ft³	Percent free gas space
1	1125	10	76	83

*Harshaw Chemical Co. "Tellerette" bulletin, and Dr. A. J. Teller.

Table 20.12 Super Intalox® Saddles: Ceramic†.

Size no. designation	Approximate number/ft³	Weight‡ lb/ft³	Surface area ft²/ft³	Percent free gas space
1	1,500	35	76	77
2	180	38	32	75
Plastic				
1	1,500	5.85	63	90
2	180	3.75	33	93
3	42	3.00	27	94

By permission, Norton Chemical Process Products Corp., Bull. Si-72 and Bull. PTP-1; other manufacturer's data are equivalent.

†Also available in polypropylene (including glass reinforced); high-density polyethylene, rigid PVC, fluorinated vinyls.

‡Weights for polypropylene; others are times PP: 1.03 for high-density polyethylene: 1.54 for PVC: 1.87 to 1.95 for fluorinated vinyls.

Table 20.13 Hy-Pak™* metal packing.

Size no. designation	Approximate number/ft³	Weight,** lb/ft³	Percent free gas space
1	850	19	97
1 ½	270	–	97
2	105	14	98
3	30	13	98

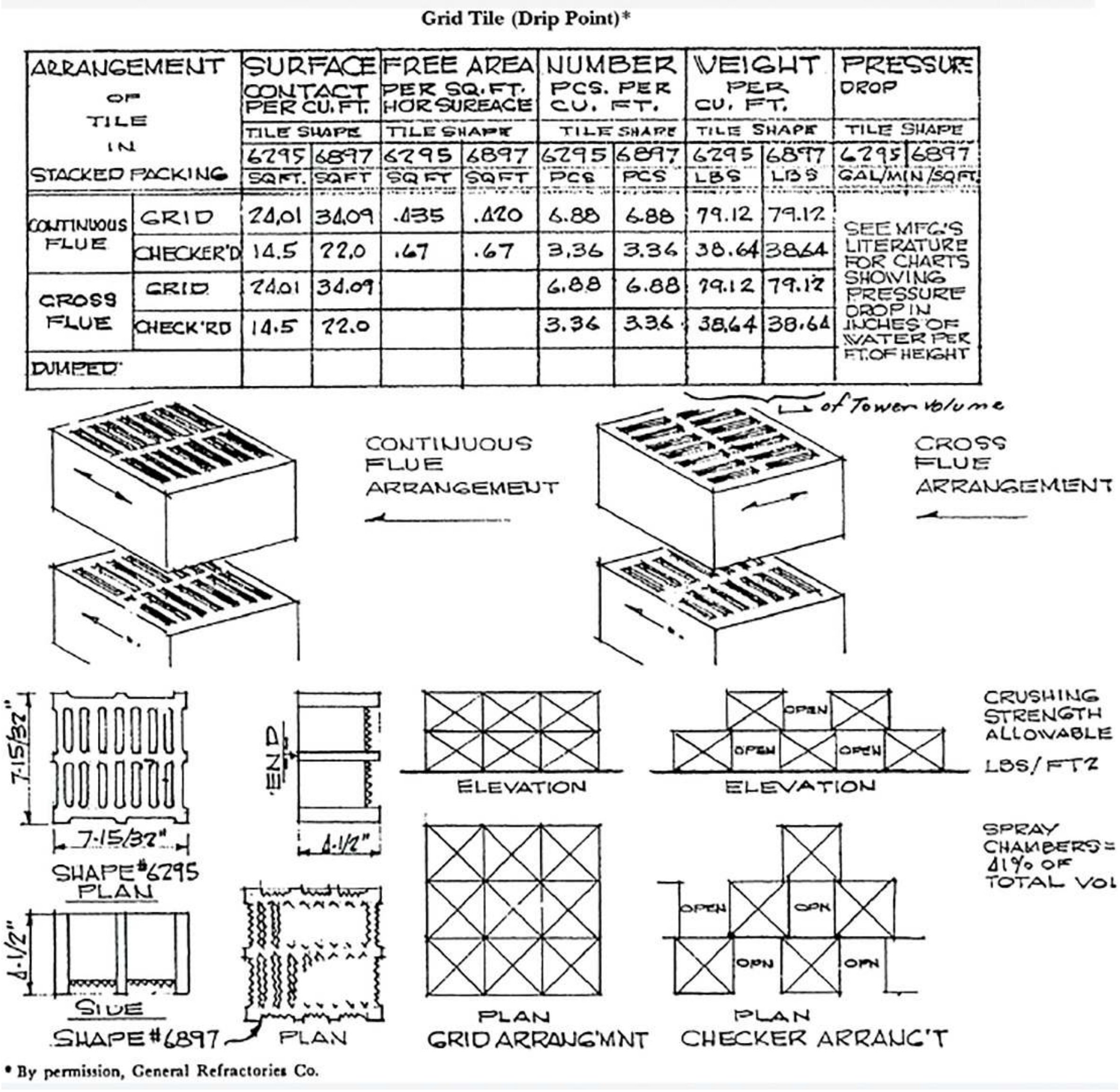
*By permission, Norton Chemical Products Corp., Bull. HY-30®, dated 9/88 reg. Trademark.
**Weight for standard gauge carbon steel, available in most other common metals.

Table 20.14 Chempak® metal packing*.

1	950	19	96

*By permission Chem-Pro Equipment Corp., Licensed from Dr. Max Levo, Bull. 702. Weight is for carbon steel.

Table 20.15 Grid tile (Drip Point)*.



Grid Tile (Drip Point)*

ARRANGEMENT OF TILE IN STACKED PACKING		SURFACE CONTACT PER CU. FT.		FREE AREA PER SQ. FT. HOR SURFACE		NUMBER PCS. PER CU. FT.		WEIGHT PER CU. FT.		PRESSURE DROP	
		TILE SHAPE		TILE SHAPE		TILE SHAPE		TILE SHAPE		TILE SHAPE	
		6295	6897	6295	6897	6295	6897	6295	6897	6295	6897
		SQ FT.	SQ FT	SQ FT	SQ FT	PCS	PCS	LBS	LBS	GAL/MIN/SQ FT.	
CONTINUOUS FLUE	GRID	24,01	34,09	.435	.420	6.88	6.88	79.12	79.12	SEE MFG'S LITERATURE FOR CHARTS SHOWING PRESSURE DROP IN INCHES OF WATER PER FT. OF HEIGHT	
	CHECKER'D	14.5	22,0	.67	.67	3,36	3.36	38.64	38.64		
CROSS FLUE	GRID	24.01	34.09			6.88	6.88	79.12	79.12		
	CHECK'RD	14.5	22.0			3.36	3.36	38.64	38.64		
DUMPED											

* By permission, General Refractories Co.

Table 20.16 Packing service application.

Packing material	General service application	Remarks
Glazed and un-glazed, Porcelain or chemical stoneware	Neutral and acid conditions except hydrofluoric, solvents. Not good in hot caustic (above 70°)	Unglazed usual type specified except special requirement of low adsorption on surface. Special ceramics available for mild caustic. Porcelain stronger and more resistant than stoneware
Carbon	Hot alkali, all acids except nitric, no oxidizing atmospheres	Stand thermal shock, low cubic weight
Plastic	Alkali, salts, aqueous, and acids depending on resin	Light weight
Steel and other light gauge metals	Hot alkali for steel, other service to suit metals	May be heavier than ceramic, more expensive

on support plates. The resulting net free area "balance" around the support grid or plate and its first two courses of packing (whether dumped or stacked) should be calculated in order to evaluate the effect on tower performance. Figures 20.1A, 20.1B, and 20.3 show a typical arrangement of several support plates.

The weight to be carried by the support plate is the sum of the weight of the packing plus the weight of the flooded liquid volume of the packing voids plus any pressure surges that might be imposed on the system. The effect of side thrust of the packing in reducing the dead packing load on the support should be ignored as it is an indeterminate figure. Normally, each support is required to support only the weights of the packed section directly above it and not those separately supported above or below it. If any intermediate supports or redistributors are not separately supported on the tower wall but rest on the packing itself, the bottom support would carry the entire tower load of packing as mentioned, plus the weights of intermediate support and redistribution plates. This is not good practice as it complicates the packing and repacking of the tower as well as putting possibly excessively heavy loads on the bottom supports.

As a general rule, packing heights per support plate should not exceed 12 ft for Raschig rings or 15–20 ft for most other packing shapes. Other types fit within these limits. Mechanical, vibrational, and thermal shock loads become important and sometimes affect the tower operation beyond these limits.

20.4 Liquid Distribution

Liquid distribution probably plays the most important part in the efficient operation of a packed tower. A good packing from the process viewpoint can be reduced in effectiveness by poor liquid distribution across the top of its upper surface or across the packing sections below any feed inlet(s) or reflux inlets. Poor distribution reduces the effective wetted packing area and promotes liquid channeling.

The final selection of the mechanism of distributing the liquid across the packing depends upon the size of the tower, type of packing (exposed surface, configuration), tendency of packing to divert liquid to tower walls, and materials of construction for distribution. Figures 20.8A–20.8L illustrate a number of available distribution types. Spray nozzles are used, but care must be taken in evaluating the fraction of the total liquid that hits the walls and never enters the packing. Full cone nozzles with spray angles that keep most of the liquid on the center portion of the packing for initial contact will perform quite well.

Spray nozzle manufacturers have spray angle data for various pressures at the nozzle inlet (pipe); see Figure 20.8L. This should be considered carefully in the distributor design, and the volume discharge per square foot of flat tower cross-section must be uniform as possible. Careful layouts of nozzle arrangements are usually required (Figure 20.8K). Maximum rate variation of ± 5–6% of average flow is necessary in design [2].

Table 20.17 Packing type application.

Packing	Application features
Raschig Rings	Earliest type, usually cheaper per unit cost, but sometimes less efficient than others. Available in widest variety of materials to fit service. Very sound structurally. Usually packed by dumping wet or dry, with larger 4–6 in. sizes sometimes hand stacked. Wall thickness varies between manufacturers, also some dimensions; available surface changes with wall thickness. Produce considerable side thrust on tower. Usually has more internal liquid channeling and directs more liquid to walls of tower. Lower efficiency
Berl Saddles	More efficient than Raschig Rings in most applications, but more costly. Packing nests together and creates "tight" spots in bed, which promotes channeling but not as much as Raschig rings. Do not produce as much side thrust; has lower HTU and unit pressure drops with higher flooding point than Raschig rings. Easier to break in bed than Raschig rings
Intalox Saddles[1] and Other Saddle-Designs	One of the most efficient packings, but more costly. Very little tendency or ability to nest and block areas of bed. Gives fairly uniform bed. Higher flooding limits and lower pressure drop than Raschig rings or Berl saddles; lower HTU values for most common systems. Easier to break in bed than Raschig rings, as ceramic
Pall Rings[2]	Lower pressure drop (less than half) than Raschig rings, also lower HTU (in some systems also lower than Berl saddles), higher flooding limit. Good liquid distribution, high capacity. Considerable side thrust on column wall. Available in metal, plastic, and ceramic
Metal Intalox®[1]Hy-Pak[1] Chempak[7]	High efficiency, low-pressure drop, reportedly good for distillations
Spiral Rings	Usually installed as stacked, taking advantage of internal whirl of gas–liquid and offering extra contact surface over Raschig ring, Lessing rings, or cross-partition rings. Available in single, double, and triple internal spiral designs. High-pressure drop. Wide variety of performance data not available
Lessing Rings	Not much performance data available, but in general slightly better than Raschig ring, pressure drop slightly higher. High side wall thrust
Cross-Partition Rings	Usually used stacked, and as first layers on support grids for smaller packing above. Pressure drop relatively low, channeling reduced for comparative stacked packings. No side wall thrust
Grid tile	Available with plan side and bottom or serrated sides and drip-point bottom. Used stacked only. Also used as support layer for dumped packings. Self-supporting, no side thrust. Pressure drop lower than most dumped packings and some stacked, lower than some ¼-in. x 1-in and ¼ -in x 2-in. wood grids, but greater than larger wood grids. Some HTU values compare with those using 1-inch Raschig rings
Teller Rosette (Tellerette)	Available in plastic, lower pressure drop, and HTU values, higher flooding limits than Raschig rings or Berl saddles. Very low units weight, low side thrust
Spraypak[3]	Compared more with tray type performance than other packing materials. Usually used in large diameter towers, above about 24-in dia., but smaller to 10-in. dia. available. Metal only
Panapak[4]	Available in metal only, compared more with tray type performance than other packing materials. About same HETP as Spraypak for available data. Used in tower 24 in. and larger. Shows some performance advantage over bubble cap trays up to 75 psia in fractionation service, but reduced advantages above this pressure or in vacuum service
Stedman Packing	Available in metal only, usually used in batch and continuous distillation in small diameter columns not exceeding 24-in. dia. High fractionation ability per unit height, best suited for laboratory work. Conical and triangular types available. Not much industrial data available

(*Continued*)

Table 20.17 Packing type application. (*Continued*)

Packing	Application features
Sulzer, Flexipac, and similar	High efficiency, generally low-pressure drop, well suited for distillation of clean systems, very low HETP
Goodloe Packing[5] and Wire Mesh Packing	Available in metal and plastic, used in large and small towers for distillation, absorption, scrubbing, liquid extraction. High efficiency, low HETP, low-pressure drop. Limited data available
Cannon Packing	Available in metal only, low-pressure drop, low HETP, flooding limit probably higher than Raschig rings. Not much literature data available. Used mostly in small laboratory or semi-plant studies
Wood Grids	Very low pressure drop, low efficiency of contact, high HETP, or HTU, best used in atmospheric towers of square or rectangular shape. Very low cost
Poly Grid[6]	Plastic packing of very low pressure drop, developed for water-air cooling tower applications

Compiled from literature

[1] Trade name of Norton Chemical Process Products Corp.

[2] Introduced by Badische Anilin and Sodafabrik, Ludwigshafen am Rhein.

[3] Trade name of Denholme Inc., Licensed by British Government.

[4] Trade name of Packed Column Corp.

[5] Trade name of Packed Column Corp.

[6] Trade name of The Fluor Products Co.

[7] Trade name of Chem-Pro Equip. Corp.

There are many other types and variations in addition to those listed, although they are usually special-purpose trays and not necessarily generally adaptable.

Good design generally considers that the streams of liquid should enter onto the top of the packing on 3- to 6-in. square centers for small towers (less than 36 in. in diameter) and should number $(D/6)^2$ streams for 36 in. and larger towers, where D is the tower inside diameter in inches [3]. When the liquid stream spacings exceed 6-in. square pitch, consideration should lean to this figure. Most manufacturers make some type of distributor giving one stream every 6 in.^{2} of tower area.

The number of irrigation or “drip-points” or entrance points per square foot of flat surface of the tower should be uniform for orifice, weir-type gravity, or pressure distributors, and need not exceed 10 points/ft^2 [2]. The uniformity must not be disturbed by support rings that support the distributor itself. The distribution must include area adjacent to the wall, and the design must not “force” more liquid at the wall where it contacts the packing. Uniformity of points of distribution to the packing surface is extremely important. The volume flow per point must be carefully calculated.

Bonilla [4] presents an excellent examination of liquid distributors in packed towers. Those with random packing or structured packing are more sensitive to poor or non-uniform distribution of liquid than tray towers. This requires that liquid and vapor are evenly distributed when they enter the packing. Often, only the liquid distributor at the top

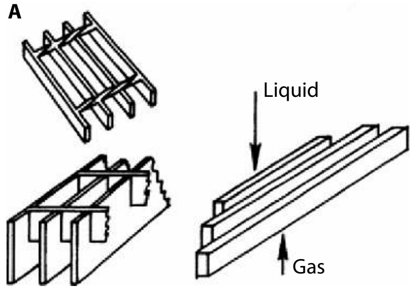
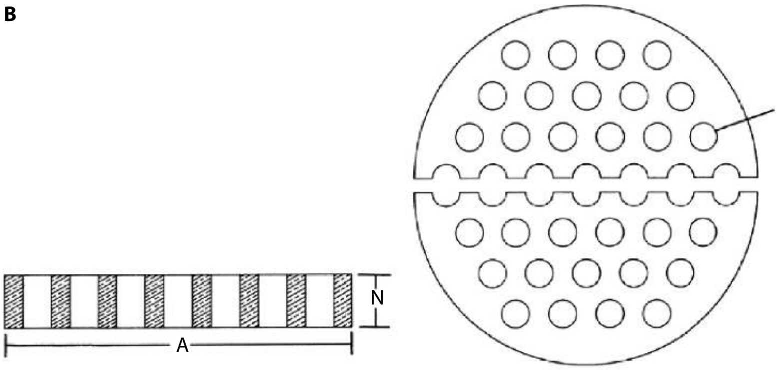


Figure 20.7A Bar-grid support. Used by permission of U.S. Stoneware Co. (now, Norton Chemical Process Products Corp.).

B



DIMENSIONS PERFORATED SUPPORT PLATE

Tower Diameter in Inches. .	12	15	18	20	24	30	36	42	48	60
Diameter of Plate "A".	11	14	17	19	23	29	35	40	45½	57½
Height of Plate "N".	1	1	1	1	1	1½	2	2½	3	4
Diameter of Holes "C".	1	1¼	1¼	1½	2	2	2½	3	3½	3½
Number of Holes	22	31	31	38	37	61	61	55	61	101
Approximate weight (lbs) . .	16	19	23	28	35	85	150	185	260	425

Figure 20.7B Perforated support plates. Used by permission of U.S. Stoneware Co. (now, Norton Chemical Process Products Corp.).

C

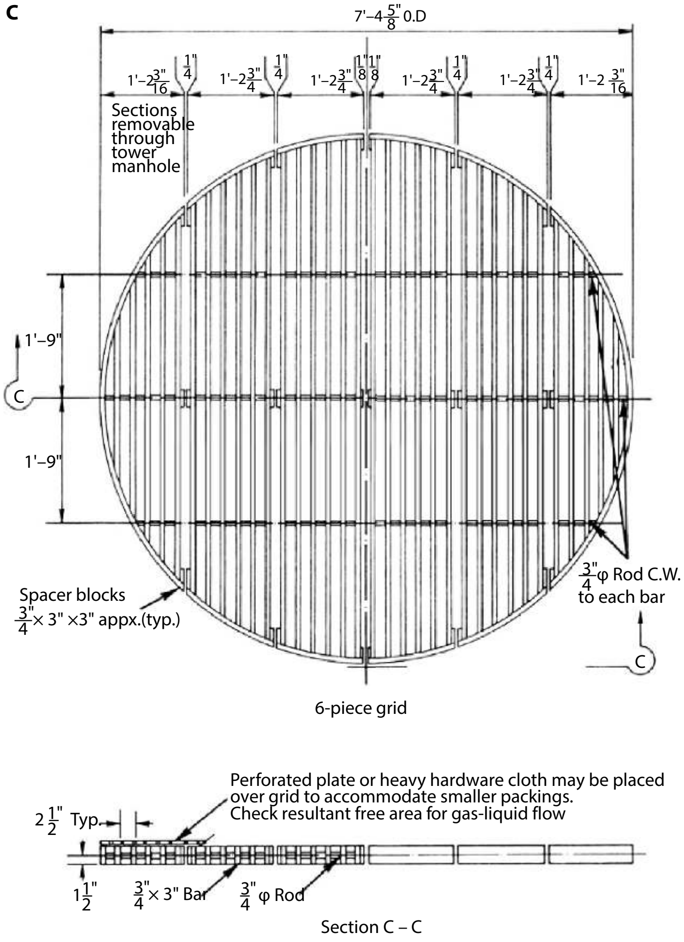
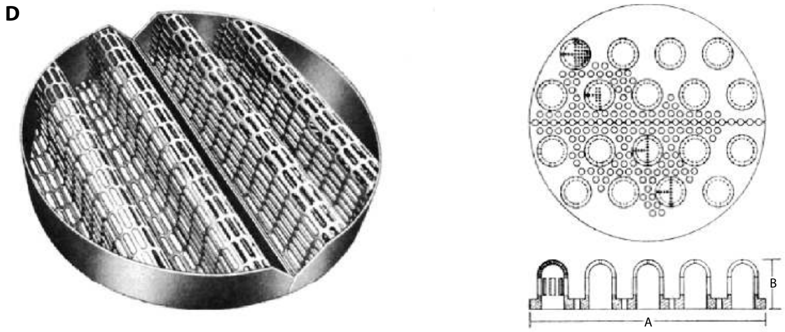


Figure 20.7C Bar-grid support plate, typical details.

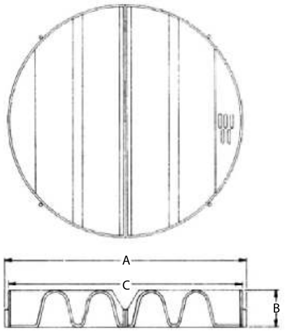


DIMENSIONS FOR SUPPORT PLATE (CHEMICAL PROCELAIN)

Tower Diam. (Inside)	Plate Diam. "A"	No. of Risers	Overall Height "B"	Per Cent of Free Area	Packing Size (Min.)	Support Ledge (Min.)‡	Approx. Net Wt. (lbs.)
12"	11.1/4"	4	5-3/4"	52.5	1/2"	3/4"	10
14"	13.1/4"	4	5-7/5"	53.0	1/2"	3/4"	14
15"	14.1/4"	4	6-7/4"	54.6	1/2"	3/4"	16
16"	15.1/4"	4	7"	56.8	1/2"	3/4"	18
18"	17.1/4"	4	7-5/16"	53.0	1/2"	3/4"	28
20"	19"	4	7-9/16"	51.4	3/4"	1"	30
24"	23"	6	7-11/16"	54.6	3/4"	1"	33
30"	29"	10	7-7/8"	55.6	3/4"	1"	75
"36"	35"	18	8"	64.2	1"	1"	93
"42"	40-1/2"	20	8-1/4"	60.2	1"	1-1/2"	180
"45"	45-1/2"	24	8-1/2"	52.9	1-1/2"	1-1/2"	292
"60"	58"	40	9"	60.6	1-1/2"	2"	400

- For these sizes use no more than 10 feet packed depth unless designed for specific load requirement.

‡ In order to avoid blocking any free area of the support plate the maximum width of the support ledge should not be appreciably greater than the minimum width shown. In prefabricated ceramic towers, 24" and 30" towers have a 1½" support ledge, the 36" tower a 1¾ ledge and the 42" and 48" towers a 2" support ledge.



MODEL 818 "GAS - INJECTION" METAL SUPPORT PLATE							
TOWER I.D.		DIA. OVER SPACERS	HEIGHT	PLATE DIA.	SUPPORT LEDGE	MIN. DIA. OF	APPROX. NET WEIGHT
INCHES	MM	A	B	C	(MIN.)	ACCESS	(LBS.)*
12	305	------	4½"	11¾"	¾"	8"	7
13¼	337	------	4½"	12¾"	¾"	9"	8
14¼	362	------	4½"	13¾"	¾"	9"	9
15¼	388	------	4½"	14¾"	¾"	10"	10
17¼	438	------	4½"	16¾"	¾"	10"	11
19¼	489	------	4½"	18¾"	1"	11"	12
21¼	540	------	4½"	20¾"	1"	12"	17
23¼	591	------	4½"	22¾"	1"	13"	19
29¼	743	------	4½"	28¾"	1"	16"	25
36	914	35½"	4½"	34¾"	1"	14"	40
42	1067	41½"	4½"	40¾"	1½"	16"	52

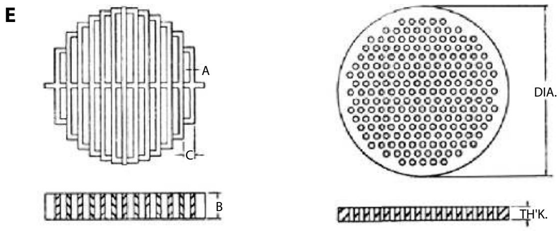
*Weights shown are for standard construction and material thickness, carbon steel.

Figure 20.7D Typical efficient metal and porcelain support plate designs for gas injection into packing. Designs also available in plastics-FRP, polypropylene, PVC, etc. (also see Figure 20.7F). Used by permission of Norton Chemical Process Products Corp.

of the packing is considered, but vapor distribution at the bottom or middle of the tower is important. The ultimate performance of a packing depends significantly on the initial distribution [4], with non-uniform distribution resulting in reduced packing efficiency, which can be expressed as a higher height equivalent to a theoretical plate/stage/tray (HETP). A structured packing with higher surface area structured packing is more sensitive to the initial liquid non-uniform distribution than a lower surface area packing.

It is important to recognize that each packing has a natural liquid distribution [4] that will develop if sufficient bed depth is available. If the distribution is poorer than the natural distribution, the system will end up with concentration gradients and higher HETP values. When/if an improvement from a poor to natural distribution occurs slowly over many feet of packing, Bonilla [4] states that adding extra packing to a bed to compensate for the initial maldistribution does not work, since the return to the natural distribution is not fast enough to compensate for the concentration gradients that have already formed.

Most liquid distributors feed onto the packing by gravity rather than being pressurized. For any given tower design, the distributor design and installation are important factors for aiding the packing to do its job. For best performance for the average distillation, the distributor should be the installed level to a tolerance of ± ⅛ to 1/16 in. and be leveled once it is in the tower. Table 20.18 [4] presents a comparison of many of the factors necessary to the



Grid bar assembly

Perforated plate

Haveg packing supports

Tower Dia.	GRID BARS No.	Size A	Size B	Size C	% Free Area	PERFORATED PLATE Dia.	Th'k	Holes No.	Holes Dia.	%Free Area
12″	7	½	2½	1½	48	13½	1	33	¾	6
15″	9	¾	2½	1½	30	13⅝	1¼	55	¾	10
18″	9	¾	2½	1½	34	16½	1½	65	¾	8
20″	9	¾	2½	1½	37	19½	1⅝	95	¾	9
2″	13	¾	3½	1½	39	23¼	1¾	134	¾	9
2′-6″	15	¾	3½	1½	36	29¼	2	188	¾	7
3″	19	¾	3½	1½	38	35	2	299	¾	9
3′-6″	21	¾	3½	1¾	41	41¼	2	402	¾	9
4′	27	¾	5	1¾	41	47	2¼	544	¾	10
5′	31	¾	5	1¾	45	58½	2½	843	1	19
6′	37	¾	5	1¾	46	71	3	1277	1	20
7′	43	¾	6	1¾	46	..	...	....	...	..
8′	51	¾	6	1¾	47	..	...	....	...	..
9′	57	¾	6	1¾	48	..	...	....	...	..
10′	61	¾	6	1¾	50	..	...	....	...	..

Note: All dimensions are in inches unless otherwise noted.
perforated plate free area should be increased when possible.

Figure 20.7E Reinforced plastic support plate. Used by permission of Haveg Corp. Bull. F-7. Note: All dimensions are in inches unless otherwise noted. Perforated plate free area should be increased when possible.

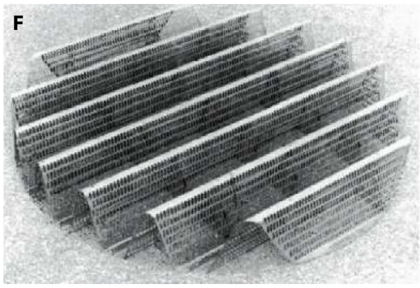


Figure 20.7F Support plate for high gas/vapor flow (also see Figure 20.7D). Used by permission of Jaeger Products Inc.

selection and design of types and styles of distributors. Selection of types depends on liquid rate, turndown requirements, plugging or fouling tendencies, feed flow, and many other considerations [4].

20.5 Packing Installation

Preparing stacked packing is a hand operation and rather costly. It is avoided where possible except for the initial layers on supports. Liquid distributed on a stacked packing usually flows straight down through the packing that is immediately adjacent to the point of contact. There is very little horizontal liquid flow. Packing patterns perform differently and are illustrated in Figures 20.16A–C.

Dumped

Dumping is the most common method of installing packing. If possible, the tower should be filled with water (for ceramic packings) after installation of the bottom support arrangement, and the loose packing floated down to rest on top of the support. The fall should be as gentle as possible, since broken packing tightens the bed and increases the pressure drop.

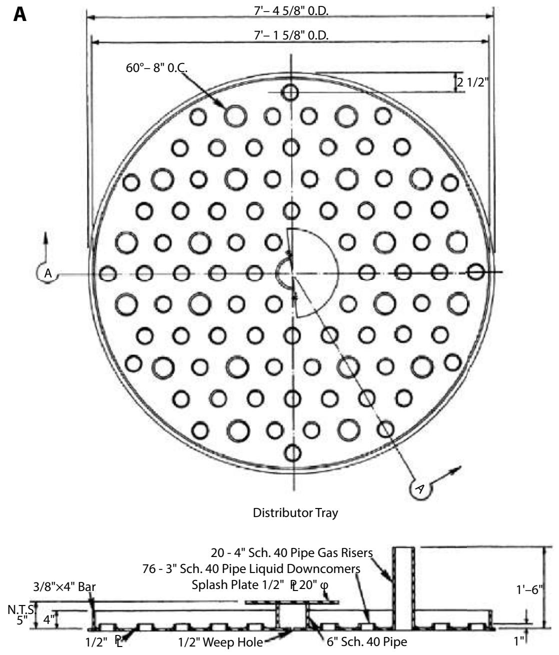


Figure 20.8A Multilevel distributor with splash plate.

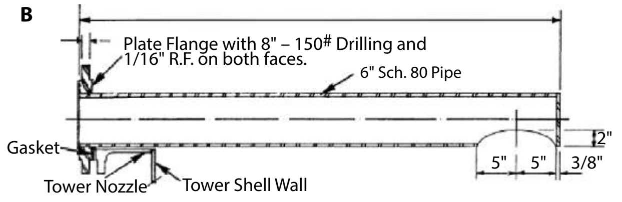


Figure 20.8B Inlet pipe for distributor.

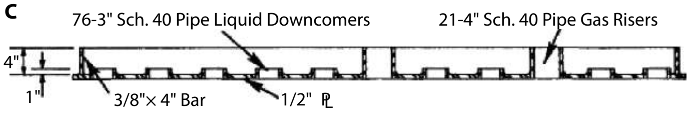
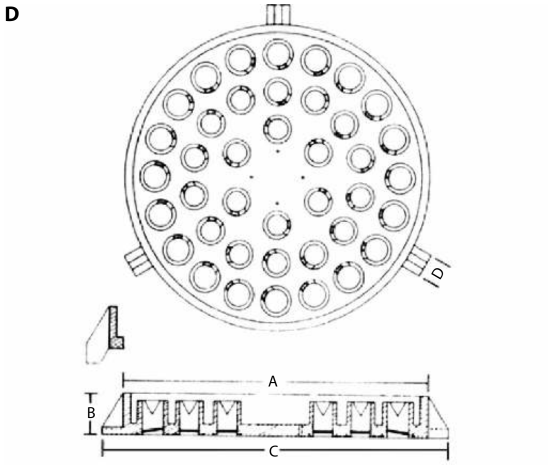


Figure 20.8C Distributor for intermediate level distribution.

One manufacturer [9] suggests loading ceramic ring type packing by using a sheet metal cone in the tower, thereby causing the rings to slide off the cone and fill to the edges first. The cone is lifted as the rings are floated or dropped in. This technique is believed to reduce the tendency for rings to channel liquid to the tower was. Saddle-type packing does not require the use of the cone.

In packing a tower dry, high hydrostatic heads are thus avoided on joint connections; however, extra care must be taken to avoid ceramic breakage. The packing will probably settle after installation, but it should not be pressed or tamped into place, as this will cause extra breakage. Furthermore, the packing should not be allowed to fall more than two feet to the bed surface; it should be dumped at random to avoid any pattern developing. The dry packed tower will be denser than the wet packed. The pressure drop for dry packed beds can be as much as 50–60% greater than for wet packed. Dry packing is not the preferred method due to significant settling that will occur.

Metal and plastic packing can be dumped into the tower when dry; however, reasonable care must be exercised and the drop should not be more than 10–15 ft, using a chute with a cloth sock (open) on the end to cushion the impact. Repacking a tower will usually change the pressure drop; for small 8-in. diameter units, this may be as much as 100%. In larger columns (24-in. diameter and higher), this variation is appreciable: about 50% or less. The various packings have characteristic ways of distributing the liquid throughout the bed. A general summary is given in Table 20.19.



DIMENSION FOR "WEIR FLOW" DISTRIBUTOR (CERAMIC)

Tower Dia.	Plate Dia. "A"	Overall Dia. "B"	Overall Dia. "C"	Width of Lugs "D"	No. of Weirs	Flow (Gallons Per Minute)	Approx. Net Weight
24"	20"	23"	5¾"	3"	8	16	55
30"	25"	29"	5¼"	3"	18	36	70
36"	30"	35"	5¼"	3"	18	36	120
42"	35"	41"	5½"	3"	36	72	150
48"	40"	46"	5½"	3"	40	80	190
60"	50"	58"	6"	4"	48	96	270

*Raised lugs standard on this size only.

Figure 20.8D Weir-flow distributor. Used by permission of U.S. Stoneware Co., Bull. TA-40 (now, Norton Chemical Process Products Corp.).

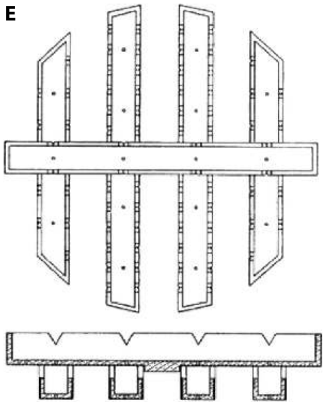


Figure 20.8E Trough liquid distributor. Used by permission of U.S. Stoneware Co. (now, Norton Chemical Process Products Corp.).

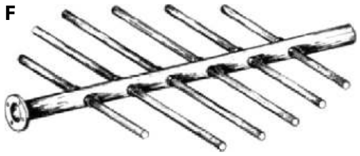
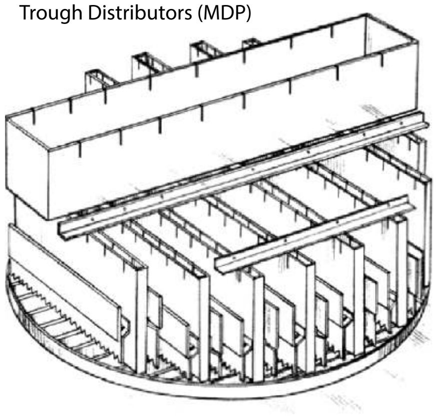


Figure 20.8F Typical design inlet liquid distributor using holes (orifices) on underside of distributor pipes. Note: Number of side pipes adjusted to provide uniform entry per square foot of water cross sectional area. Holes at wall should be spaced on the same basis.

G

Koch/Sulzer
Trough Distributors (MDP)



This distributor is widely used for structured packing applications in medium-to-large-diameter columns. The header above the lateral arrangement provides for cost-effective sectionalized designs without the need of gaskets. Elevated holes in both the headers and laterals provide superb fouling resistance, As a result of multiple drip points (MDP), a cutain of liquid falls perpendecular to the top of the packing.

In a typical desesign, the header box my have holes in the floor or, for fouling services, holes in metering boxes that are welded insides the header. The header is levealed independently by lugs attached to the laterals. Laterals have holes punched in the sides above the floor of the lateral. The MDP baffle arrangement directs the liquid onto the packing with a typical lateral hole density of 10 points/ft^2; Turndown capabilities: typically 2.5:1. A two-stage design can be provided for higher turndown requirements.

Overall distributor height is approximately 24". The distributor is leveled from a supporting grid, which normally rests on the top of the packed bed. The vessel manway is usually located at or above the header elevation.

*U.S. Patent: 4,816,191
Canadian Patient: 1,261,249
**See note

NOTE:**SULZE® is the registered trademark of Sulzer Brathers Ltd, Winterthur, Switzerland. Koch Engineering Company, Inc. is the exclusive lincesee and distributor of packing. infermals static mixers, mist elimitnators and other quality process equipment for the chemical, petrochemical and petroleum industries, bearing the SULZER® trademark in the U.S. Canada and Mexico.

Figure 20.8G Koch/Sulzer® trough distributor (MDP), for medium to large diameter columns for structured packing (patented), one of several designs available. Used by permission of Koch Engineering Co., Inc. Bull. KS-6.

H

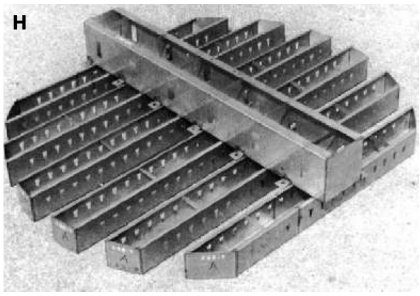


Figure 20.8H Orifice pan liquid distributor with round or rectangular chimneys and a flat floor sealed to vessel support ring. For a large vessel, it can be sectioned to pass through a manway. Used by permission of Jaeger Products, Inc., Bul.-1100.

I

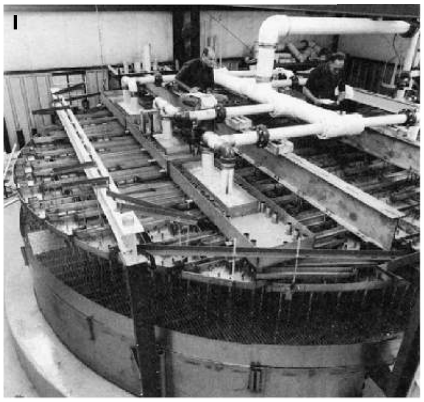


Figure 20.8I Multipan MTS-109 two-stage liquid distributor. Used by permission of Nutter Engineering, Harsco Corp., Bull. CN-4.

J

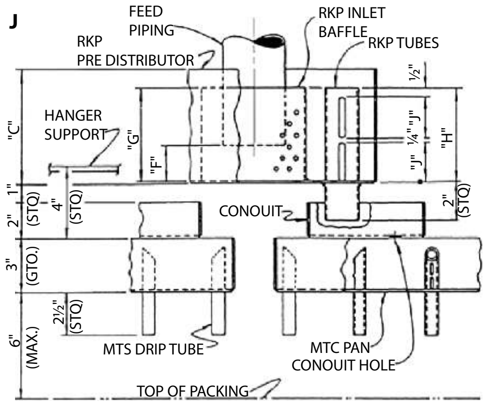


Figure 20.8J MTS-109 Multipan two-stage liquid distributor for optimum liquid distributor for uniform flow; for random and structured packings for low to moderate liquid rates, less than 5 gpm/ft^2. Also used for redistributor. Used by permission of Nutter Engineering, Harsco Corp. Bull. T1-1, under license from the Dow Chemical Co., protected by U.S. Patents Nos. 4, 472, 325; 4,808,350; 5,013,491.

K

312" diameter spray distributor for refinery vacuum tower

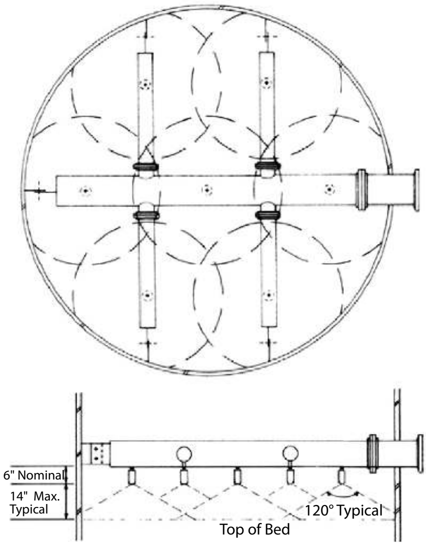


Figure 20.8K Spray nozzle distributor using full-cone wide-angle spray nozzles (depends on tower diameter). Not recommended for distillation applications, where the point-type distributors provide higher packing efficiency. Used by permission of Nutter Engineering, Harsco Corp., Bull. T1-1.

L

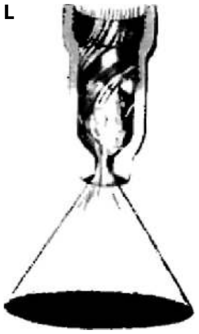


Figure 20.8L Full cone spray nozzle as used for liquid distribution. Used by permission of Spraying Systems, Co., Cat. 55.

Table 20.18 Guidelines for selection of liquid distributors.

Type of distributor						
Factor	**PAN (RTD)**	**POH (gravity)**	**POH (pressure)**	**NTD**	**SNH**	**VND**
Uniformity	VG	VG	F	VG	P	P
Solids handling	P	P	F	G	G	VG
Turndown	G	G	G	G	P	P
Ease of installation	F	F	G	F	VG	G
Ease of leveling	F	G	VG	F	VG	F
As redistributor	G[1,2]	No[2]	No	No[2]	No	No[2]
Height requirement	M	L	VL	H	M	H
Cost	H	M	L	H	L	M
Residence time	H	L	L	M	L	M
Suitable for large diameters (>10 ft)	P	G	G	G	VG	G
Leakage potential	H[3]	No	No	No	No	No
At high vapor rates	P	G	G	G	VG	G
At high liquid rates	G	G	G	G	VG	G
At low liquid rates	G	VG	G	VG	P	P
For high-purity fractionation	VG	VG	P	VG	P	P
Heat transfer	G	G	F	G	VG	G
Liquid feed handing	Yes[1]	No	No	Yes	No	Yes
Flashing feed handing	Yes[1,2]	No[2]	No	No[2]	No	No[2]

Key: VG = very good; G = good; F = fair; P = poor; H = high; M = moderate; L = low; and VL = very low. "No" means that the distributor cannot be used for the particular factor.

[1] No significant liquid mixing provided, however.

[2] Very good if used in conjunction with a chimney tray.

[3] Leakfree if seal-welded.

Reproduced by permission of The American Institute of Chemical Engineers; Bonilla, J.A., *Chem Eng. Prog.* Vol. 89, No. 3 (1993) © p. 47; all rights reserved.

20.5.1 Packing Selection and Performance

Guide Lines: Trays versus Packings

Kister *et al.* [5] prepared one of the few comprehensive distillation studies for the selection of valve and sieve trays compared to random or structured packing. This reference is based on a more comprehensive evaluation of accumulated data by the same authors [6]. Many separate studies have been conducted for trays [7] including bubble caps as well as various packings, but few, if any, attempt to establish similar conditions in order to make a viable comparison as is attempted in References 5 and 6. There are two main considerations related to capacity and separation [5]:

1. An optimal tray design is one that balances tray and downcomer area so that neither prematurely restricts capacity. Weir height, weir geometry, clearance under the downcomer, and fractional hole area are set to maximize efficiency and capacity.

Table 20.19 Liquid distribution patterns in packed columns (Data in 6-in., 12-in., and 24-in. Dia. Towers).

Packing	Tower dia. in.	Type liquid feed	Percent liquid distribution in inner ¾ tower area	
			2 ft from top	8 ft from top
1/2 in. Raschig Rings	6	Center Point	40	40
	12	Center Point	90	78 (4 ft down)
	12	4 Point	70	65 (4 ft down)
1-in. Raschig Rings	2	Center Point	95	70 (4 ft down)
	2	4 Point	70	60 (4 ft down)
½-in. Berl Saddles	6	Center Point	60	50
	12	Center Point	95	88 (4 ft down)
	12	4 Point	70	60 (4 ft down)
1-in. Berl Saddles	12	Center Point	85	65 (4 ft down)
	12	4 Point	75	55 (4 ft down)
1-in Lessing Rings	12	Center Point	90	55
	12	4 Point	70	55
	12	19 Point	70	50
	24	Center Point	100	90
	24	12 Point	80	70
½-in. Glass Rings	6	Center Point	90	70 (7 ft down)
	6	Center Point	83	75 (7 ft down) 3 ft/sec, air
	6	4 Point	75	70 (7 ft down)
	6	4 Point	75	75 (7 ft down) 3 ft/sec, air
½-in. Spheres	6	Center Point	83	85 (4 ft down)
	6	Center Point	90	90 (4 ft down, 2.5 ft/sec, air)

*Compiled from M. Leva. *Tower Packings and Packed Tower Design*, 2nd Ed., U.S. Stoneware Co. (now, Norton Chemical Process Product Corp.) (1953), Ref. 40, by permission.

2. Optimal packing design configures distributors, supports, and bed heights to avoid loss of efficiency to maldistribution and premature capacity restriction [5].

The four main differences between bubble caps and packings are [5]:

1. Differences between the capacity and efficiency of an optimal tray and an optimal packed tower design.
2. Deviations from optimal design of trays, packings, and other tower internals, e.g., distributors and baffles.
3. Unique system characteristics and special design features, e.g., corrosion, foaming, chemical reaction, and fouling, and designs to overcome such problems.
4. Capacity and separation gains due to lower pressure drop of packing. Pressure drop of packing is typically three to five times lower than that of trays.

Due to the need to use case-by-case analysis, the Kister studies [5, 6] focused on item 1. The data evaluated came from published reports by Fractionation Research Inc. (FRI) and Separation Research Program (SRP) at the University of Texas, taken from commercial size equipment rather than laboratory research columns. The FRI data include No. 2 and No. 2.5 Nutter random rings packing, and Norton's Intalox® 2T structured packing. Each considered currently state-of-the-art or close to it, while the sieve and valve trays of FRI's latest designs, plus Nutter's proprietary valve trays, all using 24-in. tray spacing.

To allow for the vertical height required for packed columns for tower distributors and redistributors typically 10%–20% of the vertical packed height (10% for 2-in. random packing and 20% for structured packing) [5] and in tray towers, the height required by additional trays, the analysis indicated:

$$\text{Practical packing HETP, HETP}_{\text{packing,Practical}} = (\text{m})(\text{HETP})_{(\text{Test packing})} \tag{20.1}$$

$$\text{For practical trays HETP, (HETP)}_{\text{Tray,practical}} = 97.5(\text{S/E}) \tag{20.2}$$

where

m = factor higher than test HETP; = 1.1 for 2-in. random packing
= 1.2 for structured packing used
S = tray spacing, in.
E = overall column efficiency, %

Capacity factor:

$$C_S = U_G\sqrt{\rho_G/(\rho_L - \rho_G)} \tag{20.3}$$

Tray spacing [5]:

$$C_{S,\text{flood}}\ \alpha\ S^{0.5} \tag{20.4}$$

Correlating liquid rate and pressure, see Figure 20.9.

$$FP = L/V\sqrt{\rho_G/\rho_L} \tag{20.5}$$

$$\text{Capacity parameter} = C_S F_P^{0.5} \nu^{0.05} \tag{20.6}$$

where

C_S = capacity factor, ft/s, based on tower superficial area
E = overall column efficiency, trayed column, %
FP = flow parameter, dimensionless
HETP = height equivalent of a theoretical plate, in.
L = liquid flow rate, lb/h-ft^2 of cross-sectional area
m = constant, allowing for vertical tower height consumed by distribution/redistribution equipment
S = tray spacing, in.
U = velocity, ft/s based on tower superficial area
U_G = superficial velocity based on cross-section area of empty column, ft/s
V = vapor flow rate, lb/h-ft^2 of cross-sectional area
ρ = density, lb/ft^3
Fp = packing factor, empirical
N = kinematic viscosity, liquid, centistokes (kinematic viscosity = viscosity, centipoise/specific gravity) (not density)

Subscripts:

G = gas
L = liquid

Figures 20.9 and 20.17 plot flood capacity versus flow parameter. The FP values of 0.4–0. 7 are estimated by Kister *et al.* [5] in absence of data. The plots show that for low and moderate pressures, the flood capacity factor versus FP correlates the effects of liquid rate and pressure on the optimized tray capacity [5]. At higher pressures, an additional effect of pressure on capacity shows a decline of optimized tray capacity.

Figures 20.10 and 20.18 plot HETP versus FP for optimized trays at 24-in. spacing, and No. 2 and No. 2.5 Nutter Rings and Intalox® 2T structured packing.

In summary, we quote the conclusions of Kister *et al.* [5]:

"Comparing trays at 24-in. tray spacing with a state-of-the-art 2–2.5-in. (nominal) random packing, and with a state-of-the-art structured packing of 67 ft^2/ft^3 specific surface area, all optimally designed, we found that:

At FPs of $\cong 0.02 - \cong 0.1$:

- The trays and the random packing have much the same efficiency and capacity.
- The structured packing efficiency is about 50% higher than either the trays or the random packing.
- As FP increases from 0.02 to 0.1, the capacity advantage of the structured packing (over the trays or over the random packing) declines to 0 from 30–40%.

At FPs of 0.1–0.3:

- The trays and the random packing have much the same efficiency and capacity.
- The structured packing has much the same capacity as the trays and the random packing.
- As FP increases from 0.1 to 0.3, the efficiency advantage of the structured packing over the random packing and over the trays declines to about 20% from about 50%.

At FPs of 0.3–0.5:

- Efficiency and capacity for the trays, the random packing, and the structured packing decline with a rise in flow parameter.
- The capacity and efficiency decline are steepest in the structured packing and shallowest in the random packing.

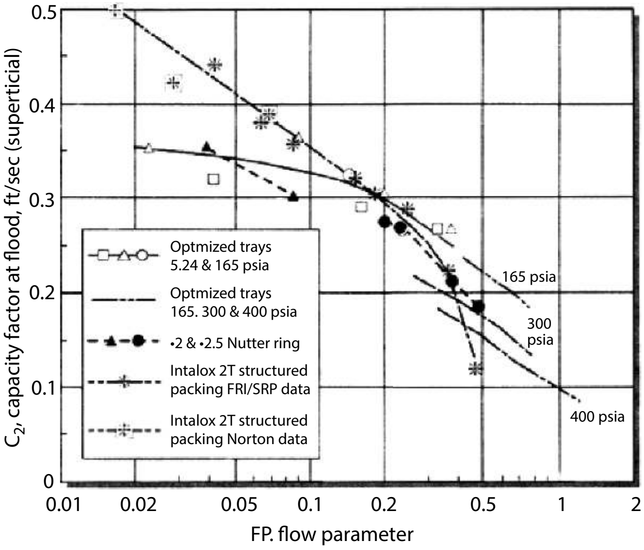


Figure 20.9 Overall comparison of capacity at flood for 24-in. tray spacing with random packing. Reproduced with permission of the American Institute of Chemical Engineers, Kister, H. Z., Larson, K. F., and T. Yanagi, Chemical Engineering Progress, V. 90, No. 2, (1994), p. 23. All rights reserved.

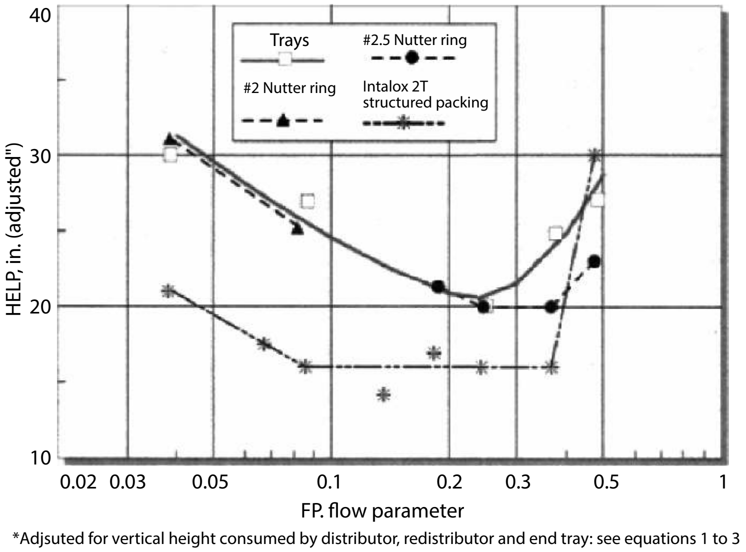


Figure 20.10 Overall comparison of efficiency for "state-of-the-art" random and structured packing with trays at 24-in. spacing. Reproduced with permission of the American Institute of Chemical Engineers, Kister, H. Z., Larson, K. F., and T. Yanagi, Chemical Engineering Progress, V. 90, No. 2, (1994), p. 23. All rights reserved.

- At an FP of 0.5 and 400 psia, the random packing appears to have the highest capacity and efficiency and the structured packing the least.

The above results are based on data obtained for optimized designs and under ideal test conditions. To translate our findings to the real world, one must factor in liquid and vapor maldistribution, which is far more detrimental to the efficiency of packings than of trays. In addition, one also must account for poor optimization or restrictive internals, which are far more detrimental to the capacity of trays than packings."

Chen [8] highlights the long-term growth of the use of bubble cap trays, valve, and sieve trays, followed by the increased popularity of packed columns accompanied by the development of random and structured packings. There are some applications in chemical/ petrochemical/petroleum/gas treating processes where one type of contacting device performs better and is more economical than others. Chen [8] points out:

1. A typical tray has opening area ranging 8% to 15% of the tower cross-section area.
2. A typical packed tower design has more than 50% of open tower cross-section, with the void fraction of a packed tower being higher at around 90% of tower volume, resulting in the following:
   a. **Pressure drop per theoretical stage**—Packed towers usually result in lower pressure drop per theoretical stage than trays. Trays often show ΔP of 3 to 8 mm Hg per theoretical stage, with packing having about 1 to 2 mm Hg for random packing and 0.01 to 0.8 mm Hg for structured packing. For high-pressure systems, the difference may not be significant, but for atmospheric and below atmospheric pressures, it can be quite significant.
   b. **Liquid hold-up**—Trays usually hold up 8 to 12% of tower volume, compared to 1 to 6% for packed towers. This can be significant for systems involving thermal degradation and that need very short residence times, which also aids in sharp separations.
   c. **Liquid/vapor ratios**—Trays are designed for low liquid/vapor ratios, while packed towers are operated from low to high liquid/vapor ratios (often in absorbers and scrubbers). Low L/V ratios are usually associated with distillation, with (approx.) less than 10 gallons of process liquid/(min)(ft^2) (tower cross-section area).
   d. **Liquid cooling**—This is usually better handled in tray towers, and it is easier to draw off liquid from trays for removal from the system, or for external recirculation.

- **e. Foaming systems**—The contact surfaces of packings promote film action compared to the droplets from trays for mass and heat transfer: the packings tend to be more resistant to entrainment and induce less foaming.
- **f. Corrosion**—Corrosion problems with some fluid systems dealt with more easily and at lower cost by corrosion resistant packings than fabricated trays.
- **g. Solids and slurries**—Trays will handle solids and slurry systems better than packings; and if solids build-up does occur, washing/flushing treatment will usually "wash" or dissolve the solids attached to the trays more easily than by attempting a thorough cleaning of packing.
- **h. Costs**—Other than special needs requiring one contacting mechanism or another, small-diameter columns of 18-in. diameter or less can be assembled less expensively as packed towers. For some applications of larger diameter columns, the packed tower may still provide the less expensive choice. A cost analysis comparison would determine benefits for individual situations.

Chen [8] recommends the following guidelines for the design of the important distributors of liquid (still must pass vapor):

Pan-type distributor: Plate with drilled/punched holes for liquid downflow and vapor risers.

Vapor riser: 15 to 45% of tower cross-section area, round risers usually 4 or 6 in. in diameter, although the round design usually has less free area than a rectangular design. Usual standard height is 6 in.; however, any height can be used as long as it is well above the liquid height on the tray. The pressure drop through the vapor riser should be low:

$$\Delta P = 0.46\left[D_v/D_l\right][V/A]^2, \text{ in. of liquid} \tag{20.7}$$

Liquid orifices on tray pan: usually at least 10 orifices per square foot of tower area. The orifice diameter can be determined [8] by

$$d = 0.23\ [Q/(K)\ (H)^{0.5}]^{0.5} \tag{20.8}$$

where

ΔP = pressure drop through risers, in. of liquid
V = vapor rate, ft^3/s
A = riser area. ft^2
D_v = vapor density, lb/ft^3
D_l = liquid density, lb/ft^3
d = orifice diameter, in.
Q = liquid flow rate, gal/min
H = differential head at orifice, ft of liquid
K = discharge coefficient, for punched holes = 0.707

Liquid design height is usually one-half of the riser height. At minimum rates, the depth on the tray can vary from ½ in. to about 1 in. below top of riser height for maximum rates. The minimum orifice diameter is recommended at ¼ in. diameter to overcome miscellaneous plugging of the holes [8]. Experience indicates that the holes really should be ⅜ in. to avoid industrial plugging problems.

Other useful distributor types have been referred to and previously illustrated. For redistribution, the vapor risers may be 12–18 in. tall, and with protective "hats" to prevent liquid dropping from the tray/section above. The space between the cover "hat" on the riser and the bed above should be 18 to 12 in. minimum to allow for proper vapor redistribution entering the packed section above. The importance of a horizontally level distributor cannot be overemphasized.

Any of the available packings can usually be exchanged with another, the differences being in efficiency of contact, expressed as height of transfer unit (HTU), height equivalent to a theoretical plate/stage (HETP) or Kg_a, and pressure drop for the particular packing-fluid system. Therefore, system data are very important when selecting a packing.

When not available, an effort should be made to find an analogous system in terms of process type, fluids, physical properties, pressure and temperature conditions, etc. If this is not possible, then the best judgment of the designer must be used.

Eckert [9] provides some basic guidelines to good packing selection over a range of system performance requirements. Kunesh [10] illustrates the often-determined pressure drop advantage of random packed towers over the usual valve tray. See Figure 20.11 [10].

For a preliminary guide and broad comparison of packings versus various distillation tray types, Table 20.20 is helpful. Although the table includes a list of the more prominent manufacturers of trays and packing materials, it is not all inclusive. Table 20.21 [11] compares trays, random and structured packing, and HETP, where C-Factor

$$C = V_s\left[D_v/(D_l - D_v)\right]^{0.5}, \text{ft/s} \tag{20.9}$$

Souders-Brown C-Factor

$$= V_s\left[\rho_v/(\rho_l - \rho_v)\right]^{0.5} \tag{20.10}$$

and F-Factor

$$F = V_s\sqrt{D_V} = V_s\sqrt{\rho_V}, (\text{ft/s})\left(\text{lb/ft}^3\right)^{0.5} \tag{20.11}$$

where

V_s = superficial vapor velocity, ft/s. (across tower cross-section)
D_v = ρ_v = vapor density, lb/ft^3
D_l = ρ_l = liquid density, lb/ft^3

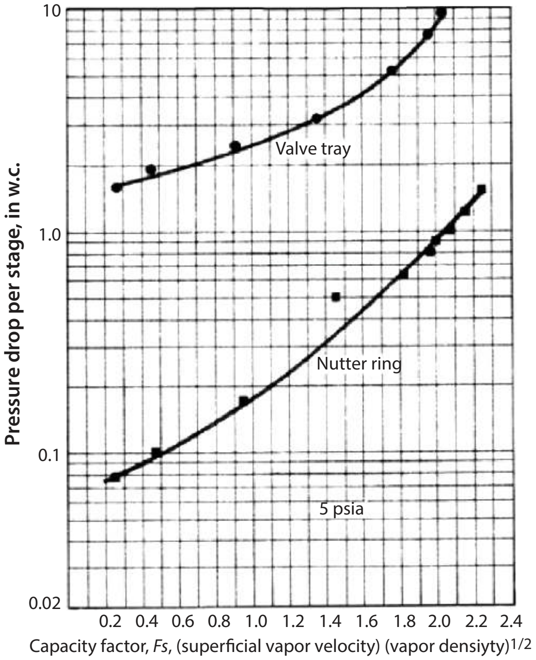


Figure 20.11 Comparison of typical valve tray and random packing showing that packing reduces pressure drop significantly. Used by permission of Kunesh, J.G., Chemical Engineering, V. 94, No. 18 (1987), p. 101, All rights reserved.

Table 20.20 Relative performance characteristics of tower packing and column trays[1].

Company	Configuration	Material of const.[2]	Relative cost[3]	Mass transfer[4]	Pressure drop	Capacity	Comments
GLITCH, INC							
Trays							
P-K trays™	Trays w/baffles	M	Mod	Med	Lo	V.Hi	
Screen Trays®	Venturi effect	M	Mod	Med	Lo	V.Hi	
Ballast®	Valve trays	M	Mod	Med	Med	Med	
Bubble Cap	Bubble cap	M	Hi	Med	Hi	Med	
Random Packing							
Ballast®	Pall rings	M, P	Lo	Med	Med	Med	
Cascade Mini-Rings	Low aspect ratio	M, P, C	Mod	V.Hi	Lo	Hi	Atmospheric-high pressure service
Structured Packing							
Gempak®A	Corrugated, perforated, lanced	M, P	Mod	V.Hi	V.Lo	V.Hi	Gempak® packings designed for high to moderate vacuum service high heat transfer service
Gempak® AT	Corrugated, perforated, lanced	M	Mod	V.Hi	V.Lo	V.Hi	
Glitsch Grid® EF-25A, EF-25AP	Lattice grid	M, P	Lo	Lo	Lo	Hi	
York Product							
Structured Packing							
Goodloe®	Knit mesh	M, P	Hi	E.Hi	Hi	Hi	
Goodloe®	Biscomponent knit mesh	P + M	Lo	E.Hi	Med	Hi	
JAEGER PRODUCTS INC.							
TRI-PACKS®							
(Hacketten® in Germany)	Hollow spherical	P	Hi	E.Hi	Lo	V.Hi	New scrubbing and stripping applications

(*Continued*)

Table 20.20 Relative performance characteristics of tower packing and column trays[1]. (*Continued*)

Company	Configuration	Material of const.[2]	Relative cost[3]	Mass transfer[4]	Pressure drop	Capacity	Comments
NOR-PAC®	Cylindrical rings/ribs	P	Lo	Hi	E.Lo	V.Hi	Replacement scrubbing &stripping applications
METAL VSP®	Hollow slotted ring	M	Mod	E.Hi	Lo	V.Hi	Used in new service. Resist damage by flow upset
TOP-PAK®	Semi-spherical	M	Lo	V.Hi	Lo	V.Hi	Large diameter packing for large columns
NOVALOX® SADDLES	Smooth, beveled, longitude, ribs	C	Mod	Med	Med	Lo	Traditional shape
BERYL® SADDLES	Smooth, beveled longitude, ribs +	C	Hi	Med	Med	Lo	Excellent mechanical strength and resistance to thermal shock and chemical attack
Structured Packing							
NOR-PAC® KOMPAKT®	Multi-layers screen cylinder	P	Hi	Hi	Lo	Hi	Ideal packing for horizontal scrubbers
Metal structural packing	Vertical sheets (New design)	M	Hi	E.Hi	E.Lo	E.Hi	
KOCH ENGINEERING							
Trays							
FLEXITRAY® 4 variations	Circular slotted valve	M	Lo	Med	Med	Med	
Random Packing							
FLEXIRING®	Slotted cylinder	M	Mod	Hi	Lo	Med	
	Internal tongues	P	Lo	Hi	Lo	Med	
K-PAC®	Modif						
	FLEXIRING® increased voids	M	Mod	Hi	V.Lo	Hi	

(*Continued*)

Table 20.20 Relative performance characteristics of tower packing and column trays[1]. (*Continued*)

Company	Configuration	Material of const.[2]	Relative cost[3]	Mass transfer[4]	Pressure drop	Capacity	Comments
FLEXISADDLE®	Saddle	P, C	Lo	Med	Med	Med	
Structured Packing							
FLEXIPAC® (Mellpak® in Europe)	Geomtr. arrngd. corrug. sheets	M, P	Hi	V.Hi	V.Lo	Hi	
SULZER (KS)		M, P	V.Hi	E.Hi	V.Lo	V.Hi	
NORTON CHEMICAL PROCESS PRODUCTS							
Trays							
Valve tray	Valve	M	Mod	Med	Med	Med	
Sieve tray	Slotted cylinder	M	Mod	Med	Med	Med	
Bubblecap	Bubblecap	M	Mod	Med	Hi	Med	
Random Packing							
IMTP®	Rib finger type	M	Mod	Hi	Lo	Hi	
HY-PAK®	Slotted cylinder	P	Mod	Med	Lo	Hi	
SNOWFLAKE™	Short cylinder ribbed	P	Lo	Hi	Lo	Hi	
Pall rings	Slotted cylinders	M	Mod	Med	Med	Med	
Pall rings	Slotted cylinders	P	Lo	Med	Med	Med	
Super INTALOX®	saddle	P	Lo	Med	Med	Med	
Saddles							
Super INTALOX®	saddle	C	Mod	Med	Med	Med	
Saddles							
Structures Packing							
INTALOX® structured packing	corrugated	M	Hi	Hi	Lo	Hi	
INTALOX® wire	woven	M	Hi	E.Hi	E.Lo	Hi	
NUTTER ENGINEERING							
Trays							

(*Continued*)

Table 20.20 Relative performance characteristics of tower packing and column trays[1]. (*Continued*)

Company	Configuration	Material of const.[2]	Relative cost[3]	Mass transfer[4]	Pressure drop	Capacity	Comments
FLOAT VALVE® tray	Rectangular valve	M	Mod	Med	Med	Med	Lateral release of vapor
							Unopposed liquid flow
V-GRID® tray	Tapered slot	M	Mod	Med	Med	Med	Fixed open, high strength
							Resistant to fouling
Random Packing							
Nutter Rings	Crimped, curved slotted strips	M	Mod/Lo	Hi	V.Lo	Hi	Superior liquid spreading F.R.I. tested efficiency
Structured Packing							
SNAP-GRID™	Slotted, snaplock shape	M	Hi	Med	E.Lo	E.Hi	"I" beam configuration High capacity, non-fouling
Montz A3™	Wire-weave corrugated	M	V.Hi	V.Hi	Lo	Hi	Highest efficiency
Montz B1™	Embossed sheet metal	M	Hi	V.Hi	V.Lo	Hi	Sinusoidal corrugations Maximum surface utilized

GENERAL COMMENTS

1. Comparisons of relative cost and performance are applicable only within same manufacturer.
2. M = Metal (Generally 304 SS. Other alloy available); P = Plastic (Wide selection); C = Ceramic.
3. Costs: Mod = Moderate or = conventional packing; Lo = conventional packing; Hi = conventional packing.
4. Mass Transfer efficiency.
5. Structured packings frequently used for high vacuum service.
6. Carbon steel and other metals sometimes available.
7. Wide range of plastic generally available.
8. Packing efficiency and capacity vary with specific application. Contact vendor for assistance in making final decision.
9. There is no intention to reflect negatively on any manufacturer's packing or trays (author note).

Used by permission; W.P. Stadig, *Chemical Processing*®, Feb. (1989), Ritman Publishing Co.

Trays are usually designed with F-factor between 0.25 and 2.0 for a turndown of 8:1. Pressure drop per theoretical stage falls between 3 and 8 mm Hg. Note that bubble cap trays are on the high end, and sieve trays are on the lower end of the range. Varying tray spacing and system efficiency changes the HETP for trays between 24 and 48 in. [8]. The C-factor is the familiar Souders and Brown capacity equation.

The number of packing sizes, types (designs), and materials of construction currently available to the designer has increased considerably. Packing selection for a specific process application requires a weighing of information and an evaluation of the closest comparable data.

Table 20.21 Typical performance characteristics comparison of tower packings and trays.

	Type of internal		
Characteristic	**Trays**	**Random packing**	**Structured packing**
Capacity			
F-factor, (ft/s) $(lb/ft^3)^{1/2}$	0.25–2.0	0.25–2.4	0.1–3.6
C-factor, ft/s	0.03–0.25	0.03–0.3	0.01–0.45
Pressure drop, mm Hg/theoretical stage	3–8	0.9–1.8	0.01–0.8
Mass transfer efficiency, HETP, in.	24–48	18–60	4–30

Reproduced by permission: Chen, G. K, *Chem. Eng.,* Mar. 5 (1984) p. 40, all rights reserved.

Overview

Kunesh [10] presents an overview of the basis for selecting random packing for a column application. To decide between trayed or packed towers—the fundamental design options—a comparative performance design and its mechanical interpretation should be completed, considering pressure drop, capacity limitations, performance efficiencies (HETP), and material/heat balances for each alternative. For one example, relating to differences in liquid distribution performance, see Reference 10.

If a packed tower is chosen, the larger packing size generally provides the greater capacity, with less pressure drop, but at the expense of lower efficiency (higher HETP) than a somewhat smaller size. Packing size is the primary variable in determining optimum performance, with packing type being an important secondary consideration. Other factors, such as column diameter and length devoted to packing, also influence overall performance. Obviously, there is a close balance here, particularly between the various design shapes (types) of the different manufacturers.

For more accurate performance data on a specific packing type/size, consult the respective manufacturers and do not rely only on the generalization charts of the published literature. These charts are continuously being improved, so they are quite useful for a good approximate design (and even final in some instances). Some competitors have packings so close in design to each other that there is little real difference between them, particularly because a reasonable “factor of safety” should be applied more specifically to packing height (when separation of components is more important) than to tower diameter (volume/mass capacity).

Fractionation Research, Inc. (FRI) [10] has found that these parameters (plus a few others) affect efficiency (HETP): system to be separated, concentration of components, absolute pressure level, column diameter, and bed length, depth, or height (the latter two primarily related to the quality of liquid/gas distribution). Kunesh [10] urges caution when selecting an efficiency prediction (HETP) that is “close to the operating conditions for an accurate/final design.” Experience suggests it may be necessary to select a final design HETP from the best available data (family of compounds, pressure of operation, and specific packing type and size) and add a factor of “safety” to suit the situation, perhaps 15–30%.

20.6 Contacting Efficiency, Expressed as Kg_a, HTU, HETP

When specific data on system are not available, and often they will not be, then close comparisons should be sought. If nothing more can be done, tabulate the relative efficiency for other systems and apply judgment to select a value.

The HETP (Height Equivalent to a Theoretical Plate (stage or plate)) is the tray spacing divided by the fractional overall tray efficiency [2]. The transfer unit concept has been useful for generalized correlations [12]. Because packed towers operate with continuously changing compositions through the packed height, the concept of HETP has been to determine the number of theoretical stages (plates) required for a given separation by the usual discrete tray-by-tray method (stepwise), and then using the height of packing equivalent to one theoretical plate, and then multiply

Table 20.22 HETP comparison of 2-in. Dia. slotted rings.

System	Top pressure, mm Hg	Bed depth, ft	Averaged HETP, in.
Acetone/water	740	5	18
		10	20
Iso-octane/toluene	740	5	24
		10	30
		20	32
	100	5	24
		10	26
		20	28
Methanol/water	740	10	21
Iso-propanol/water	740	10	24
Para/ortho xylene	100	10	33
	50	10	31
	16	10	28
Ethylbenzene/styrene	100	10	22
		20	26
Basis: Test-column diameter: 24–42 in.			

Used by permission: Chen, G. K., *Chem. Eng.* Mar. 5 (1984), p. 40, all rights reserved.

to obtain the total height of packing. This requires the use of experimentally or industrially determined HETP values for the same system (or one quite close in terms of pressure, types, or families of fluids and packing size and family type; see Figure 20.10 and Table 20.21 and later discussion). Table 20.22 [8] compares several process systems and gives the corresponding average HETP for 2-in.-diameter slotted packing rings.

20.7 Packing Size

This affects contact efficiency; usually, the smaller packing is more efficient; however, pressure drop increases.

As a general guide, use

Random Packing Size, Nominal, in.	Column diameter in.
½–⅝	6–12
⅝–1	12–18
1–1½	18–24
1½–2	24–48
2–3	36–larger

The packing sizes listed relate to ceramic (where available), metal, and plastics. Plastic random packing must be used well under its softening temperature. Due to the packing weight of the bed, the packing can compress as the temperature increases, become less effective, and thus increase the system pressure drop to the point of causing flooding. Therefore, the manufacturer should be consulted regarding upper limits of operating temperatures and corrosion resistance of each plastic. Under these conditions, it is important that the plastic packing surface be "wetted" to allow film formation.

20.8 Pressure Drop

The pressure drop (ΔP) represents the frictional and kinetic energy losses and the force exerted by the operating liquid holdup through the packed bed. At constant ΔP, there is less volumetric liquid holdup for high-density liquids,

Table 20.23 Pressure drop correlations for packed towers.

Packing		**Constants**	
Type	**Nominal size, in.**	**α**	**β**
Raschig rings Ceramic	3/8	4.70	0.41
	½	3.10	0.41
	5/8	2.35	0.26
	¾	1.34	0.26
	1	0.97	0.25
	1 ¼	0.57	0.23
	1 ½	0.39	0.23
	2	0.24	0.17
Raschig rings	5/8	1.20	0.28
Metal, 1/32 in.	1	0.42	0.21
Wall, 1/16 in.	1 ½	0.29	0.20
	2	0.23	0.135
Pall rings, metal	5/8	0.43	0.17
	1	0.15	0.16
	1½	0.08	0.15
	2	0.06	0.12
Berl saddles	½	1.2	0.21
	¾	0.62	0.17
	1	0.39	0.17
	1½	0.21	0.13
Intalox saddles	½	0.82	0.20
	¾	0.28	0.16
	1	0.31	0.16
	1½	0.14	0.14

Source: Van Winkle, M., Distillation, McGraw-Hill, Inc. (1967).

but for low-density liquids, the volume of liquid holdup can be significantly greater than for water at the same ΔP. At atmospheric pressure and in vacuum services, the static head produced by the gas is negligible. In higher pressure fractionators, where more theoretical stages are needed, the static head of vapor from the condenser to the reboiler can be considerable. In such cases, the value should be added to the calculated ΔP in determining the bottom column pressure and reboiler temperature. Pressure drop may be estimated from Figure 20.13C for given packing system properties and operating conditions [13], or may be calculated by the following equation as represented by Leva [14].

$$\Delta P = \alpha 10^{\beta L}\left(\frac{G_1^2}{\rho_G}\right) \tag{20.12}$$

The equation was originally developed for air–water systems. For other liquids, L is multiplied by the ratio of the density of water to the density of liquid. Tables 20.23 and 20.23A show ΔP correlations for different types of packing material.

ΔP is important to most column designs. Recognize that pressure drop will increase due to

1. Unsteady column operations
2. Increased liquid/vapor loads

Table 20.23A Revised packed tower pressure drop correlation constants for towers operating below flooding region.

		Nominal packing size (in.)										
Type of packing	**Mat'l**		**¼**	**⅜**	**½**	**⅝**	**¾**	**1**	**1¼**	**1⅜**	**1½**	**2**	**3**
Intalox Saddles	Ceramic	α			1.04		0.52	0.52			0.13	0.14	
		β			0.37		0.25	0.16			0.15	010	
Raschig Rings	Ceramic	α			1.96	1.31	0.82	0.53			0.31	0.23	0.18
		β			0.56	0.39	0.38	0.22			0.21	0.17	0.15
Berl Saddles	Ceramic	α			1.16		0.56	0.53			0.21	0.16	
		β			0.47		0.25	0.18			0.16	0.12	
Pall Rings	Plastic	α						0.22		0.14		0.10	
		β						0.14		0.13		0.12	
Pall Rings	Metal	α				0.43		0.15			0.08	0.06	
		β				0.17		0.15			0.16	0.12	
Raschig Rings $\frac{1}{32}$ in. Wall	Metal	α				1.20							
		β				0.28							
Raschig Rings $\frac{1}{16}$ -in. Wall	Metal	α			1.59	1.01	0.80	0.53			0.29	0.23	
		β			0.29	0.39	0.30	0.19			0.20	0.14	

By permission J. Eckert, U.S. Stoneware Co. (now, Norton Chemical Process Equipment Corp.) (1958).
Equation

$\Delta p = \alpha \times 10^{\beta L}\left(\frac{G^2}{\rho_G}\right)$ (Limited to Region Below Flooding).

Δp = Pressure Drop—in. H_2O/ft of packing
G = Gas Mass Velocity—lb/sec ft^2
L = Liquid Mass Velocity—lb/sec ft^2
ρ_G = Gas Density—lb.ft^3
α and β = Constants.

3. Breakage of ceramic packings (this can be serious)
4. Compaction/deflection of plastic packings

20.9 Materials of Construction

Careful consideration should be given to fluids, temperatures of systems, aeration, etc., likely to be encountered in service. Plastic may be attractive for the application, but its long-term heat deflection characteristics needed to be determined. With time, many plastics will deform, thereby changing the packing bed characteristics, and hence the column pressure drop. As a general rule, do not select a plastic to operate at any time within 50°F of the softening or deflection temperature of the plastic.

20.10 Particle versus Compact Preformed Structured Packings

Particle packings (random) are usually (although not always) less efficient than the pre-packaged/preformed assemblies. However, particle packings are generally more flexible in loading and have better ability to handle "dirty" fluids.

Cost of the packing and its effect on the system costs must be considered, as some packings are much more expensive than others, but produce very little improved performance. Table 20.17 presents some comparisons information. The most common packings and hence the ones with the most available data are Raschig rings, Berl saddles, and Pall Rings® (Norton Co.) or equivalent.

As a rough guide, the ratio of maximum random packing size to tower diameter is

Raschig rings	1:20 (Reference 15)
	1:8 (Reference 16)
Berl saddles	1:10
lntalox saddles:	1:8 to 1:10 (Reference 17)

The 1:8 ratio is in more common use for most packings; however, recent data indicated that Raschig rings require a larger ratio approaching 1:20.

These ratios are useful in dealing with small towers and serve as guides for somewhat larger ones. There are no guides detailing the smallest sized packing to place in a tower, but ½ in. is about the smallest ceramic in use with ¾ and l in. being the most popular. Operating and pressure drop factors will usually control this selection.

Packed towers are not limited to small units; in fact, the largest processing towers for absorption and stripping operations are probably of this type. Some units are 40 and 50 ft in diameter and use 2-in. and larger packings to packing bed heights of 20–30 ft. Other units are 5–6 ft in diameter with 60 ft of packing. Towers with a 24-in. dia. and smaller are most commonly packed because of the mechanical limitations of trays in small towers.

20.10.1 Fouling of Packing

Random and structured packings are susceptible to surface fouling due to process conditions and/or the presence of oxygen as may be related to bacterial growth. Some systems will precipitate solids or crystals from solution usually due to the temperature and concentration effects. Bravo [18] discusses air–water stripping and illustrates the effects of iron that is present in these systems on packing and fouling of distributors.

Fouled packing can significantly reduce the performance of the system checked for and should be examined when packed tower systems deteriorate. For some systems, an acid or solvent flush (or wash) will accomplish the needed thorough cleaning. The need for such maintenance should be recognized in initial system design. Also, for water-soluble systems, the use of pH adjustment, pretreating with sequestering agents or biocides, ozone treatment, and other steps specific to the system's problems [18] may be necessary or helpful.

20.11 Minimum Liquid Wetting Rates

To feed enough liquid into the tower to place a wet film of liquid effectively over all the packing, a minimum wetting rate (MWR) has been evaluated for guidance in operation and design. Morris and Jackson [16] recommended the MWR shown in Table 20.25.

A minimum liquid rate for any tower packing is used by some designers as 1,500 lb liquid/(h) (ft^2 tower) referencing to liquid of properties of water.

$$L_{min} = (MWR)\ (a_t) \tag{20.13}$$

where

L_{min} = liquid rate, ft^3/(h) (ft^2 cross-section)
MWR = value of minimum wetting rate from Table 20.24
a_t = specific surface area of packing, ft^2/ft^3

The minimum wetting rate is a function of the packing material surface (Table 20.24A) and the physical properties and the liquid involved, particularly the viscosity and the surface tension. Most plastic and some metal packings require surface treatment before they will wet uniformly or at all. Without film forming characteristics on the surface area, the contact of the liquid and vapor will be poor and performance can be expected to decline. The packing should be tested for wettability in the service before selection.

Kister [19] has evaluated Schmidt's [20] somewhat complicated equation for minimum wetting rate and proposes:

$$MWR_G = (MWR_T \text{ using Table 20.24A in gpm/ ft}^2)(60/a_t)^{0.5} \tag{20.14}$$

where

MWR_G = minimum wetting rate, gpm/ft^2, generalized for other packings using Kister's evaluation
MWR_T = minimum wetting rate, gpm/ft^2 from Table 20.24A
a_t = specific surface area of packing, ft^2/ft^3

Another expression of "reasonable minimum wetting rate" [21] is given in Table 20.24A. The surface characteristics of the packing material are important in the type of liquid film (or droplets) that flows across and around and drips off of its surface. The better the specific liquid wets the packing surface and forms a moving film, the more efficient will be the packing for distillation, absorption, etc. In general, it can be noted from the table that the surfaces that tend to wet easily have the lower minimum wetting rates. The data given in Table 20.24A do not agree too well with the recommendations of Table 20.24, however. Although there is no validation, it is believed that the information in

Table 20.24 Minimum wetting rate*.

		Liquid, ft^3/(hr) (ft^2 cross sect.)
Packing	**MWR rate**	**Packing surface area per tower volume, ft^2/ft^3**
Rings (Raschig, Lessing, etc.), thru 3-in. dia.	0.85	
Grid type (wooden, etc.) (pitch 2 in.)	0.85	
All packings larger than 3 in.	1.3	
Polished metal packings and poor wetting surfaces (some plastics, glazed porcelain, etc.)	1.3, estimate to 2.5. Preferably etch surfaces to reduce problem.	

*Complied by permission from Morris and Jackson, *Absorption Towers*, Butterworth Scientific Pub. (1953) London and Imperial Chemical Industries, Ltd.

Table 20.24A Packing wetting rates related to packing material surface.

Surface	MWR_T Gpm/ft^2	Reasonable minimum wetting rate ft^3/ft^2 hr	Reasonable minimum wetting rate m^3/m^{2}h	Materials
Unglazed ceramic	0.187	1.5	0.5	Chemical stoneware
Oxidized metal	0.27	2.2	0.7	Carbon steel, copper
Surface treated metal	0.40	3.2	1.0	Etched stainless steel
Glazed ceramic	0.80	6.4	2.0	
Glass	1.00	8.0	2.5	
Bright metal	1.20	9.6	3.0	Stainless steel, tantalum
PVC-CPVC	1.45	11.6	3.5	**
Polypropylene	1.60	12.8	4.0	**
PTFE/FEP	2.00	16.0	5.0	**

*Modified by E.E. Ludwig form Glitch literature.

**Requires proper surface wetting treatment. Important that surface wetting be tested and treatment applied if necessary.

Values based on > 43 ft^2/ft^3 specific area of packing. By permission of Mass Transfer, Inc., Bull. TP/US/B1 (1978) and Glitch Bull. No. 345.

Table 20.24A is more current and represents a more recent evaluation of available data. However, the fact that the results are not identified by packing design types suggests that further evaluation of this factor is required.

Note that when packing is changed from one material of construction to another, a new minimum wetting rate will be applied to the system.

20.12 Loading Point–Loading Region

Examination of Figure 20.12 shows the pressure drop of the packed bed with gas flow and no liquid flow as the dry curve. As liquid is added to the top of the packing, the effect on pressure drop is immediately noticeable. Note that the lower part of all the liquid rate curves parallels the slope of the "dry" bed curve; however, at a point, a noticeable change in the slope of the pressure drop curve occurs. This is attributed to liquid hold-up in the bed changing from being only a function of liquid flow to being dependent on both liquid and gas flows. Although the change seems to occur for some packings at a distinct point, it is difficult to determine this accurately for all materials and is perhaps better to consider a region—from the first point of inflection of the curve to its second. Towers are usually designed to operate with gas–liquid rates in the loading region or within about 60–80% of its upper point. As will be discussed later, the deviation from the loading point is greater for some situations due to the relative proximity of the loading to the flooding point.

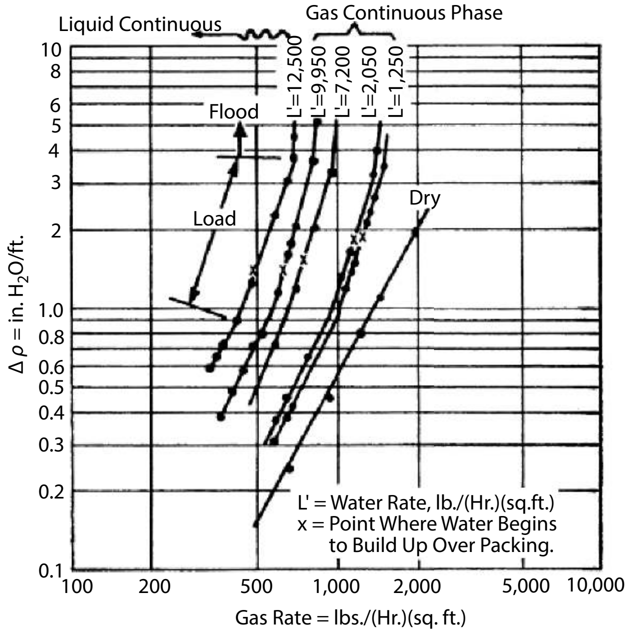
For Figure 20.13A, the loading region is centered about the 0.75 in/ft pressure drop curve, the preferred design range being 0.35 to a maximum of 1.0 in. of water/ft.

Figure 20.13D indicates that the loading region is centered about line B, which is a reasonable upper design condition.

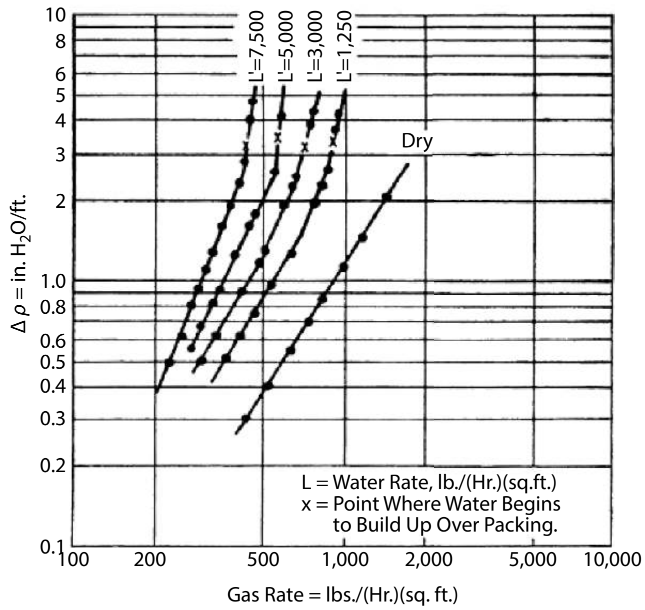
Figures 20.13B and 20.13C are the earliest generalized pressure drop correlations (GPDCs) proposed, and they have been used for many industrial plant designs. Progressively, Figures 20.13E to 20.13H are more recent correlations. These charts will be discussed in a later section.

Figure 20.13F is the most current version of the GPDC as presented by Strigle [22] to facilitate interpolation of the ordinate and pressure drop curves on the chart. The flooding and loading regions are not identified. For this chart:

$$1.\ \text{Flow parameter (FP), abscissa} = \frac{L_h}{G_h}\left[\frac{\rho_G}{\rho_L}\right]^{0.5} \tag{20.15}$$



Pressure Drop Data on 1-inch Raschig Rings



Pressure Drop Data on 1/2-inch Raschig Rings

Figure 20.12 Pressure drop flow characteristics in conventional packed towers. Reproduced by permission of the American Institute of Chemical Engineers, Sarchet, B. R., Trans. Amer. Institute of Chemical Engineers, V. 38, No. 2 (1942), p. 293. All rights reserved.

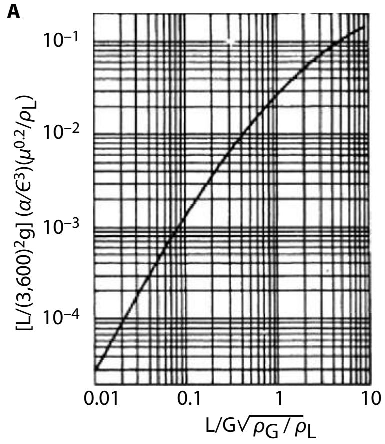


Figure 20.13A Sherwood – type correlation for flooding gas rate at a given liquid rate. Used by permission of Zenz, F. A., Chemical Engineering, Aug., (1953), p. 181. All rights reserved.

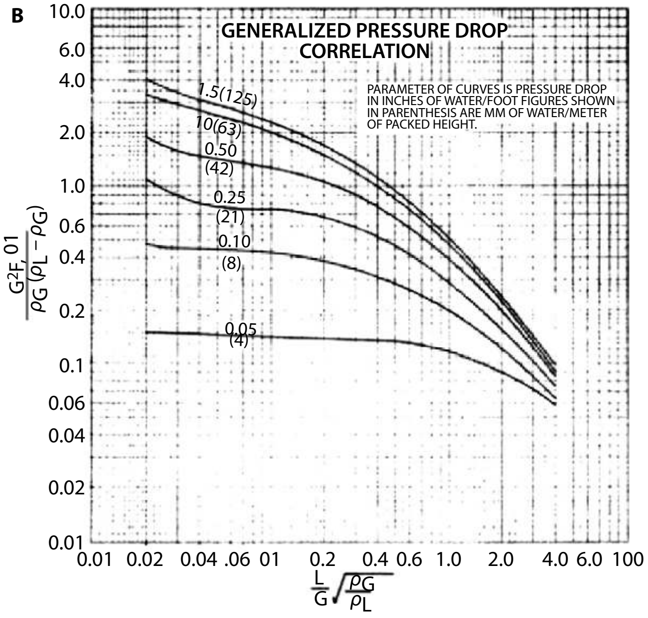


Figure 20.13B Recent generalized pressure drop correlation. Note: υ = viscosity of liquid in centipoise. Used by permission of Norton Chemical Process Products Corp.

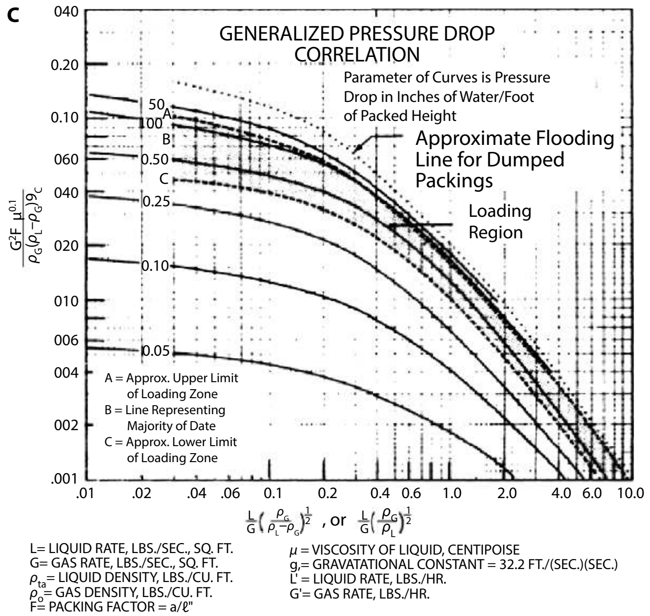


Figure 20.13C Generalized pressure drop correlation essentially equivalent to Figure 20.13B. Used by permission of Norton Chemical Process Products Corp.

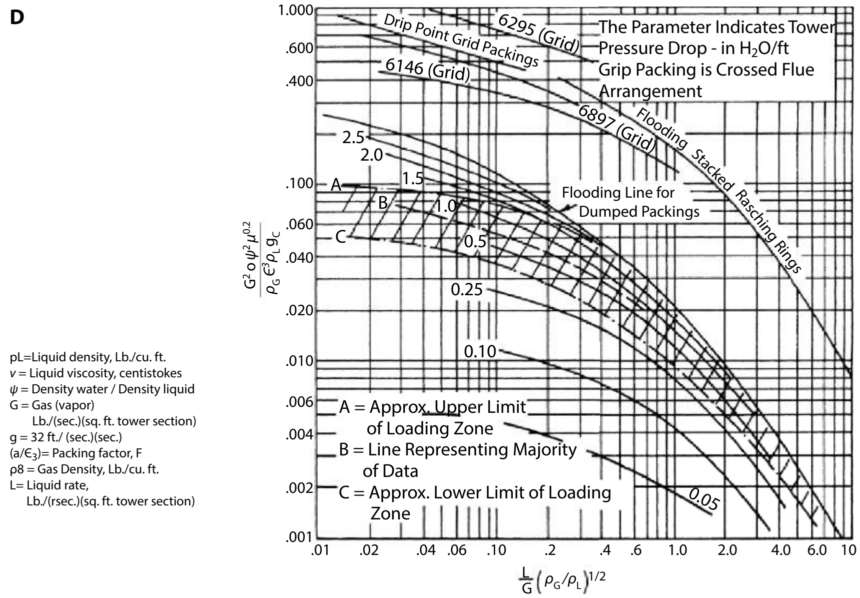
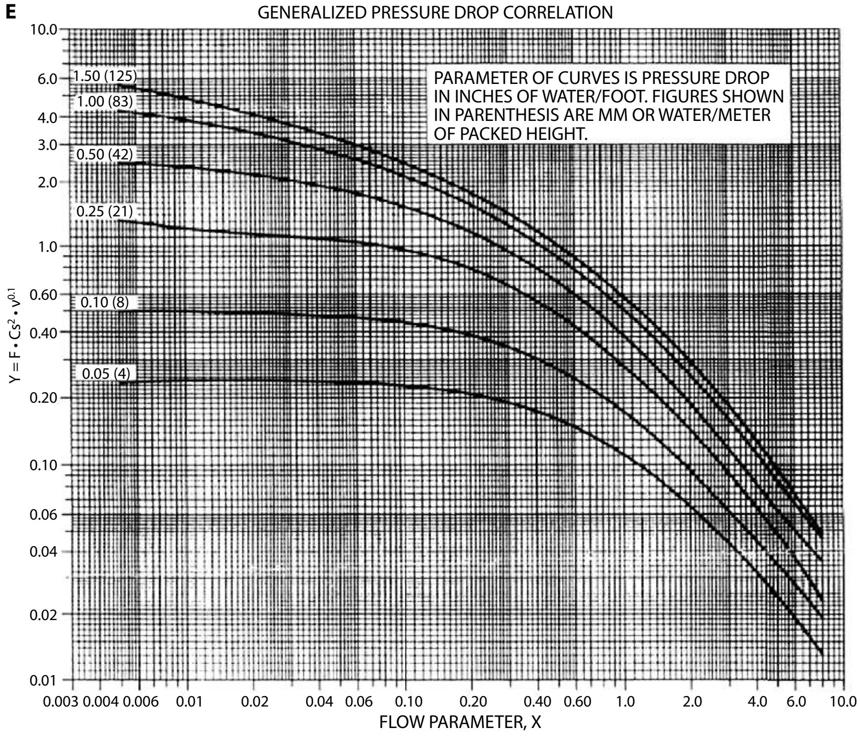


Figure 20.13D Loading, flooding and pressure drop correlation (one of earlier versions). Adapted by permission from Leva, M. Tower Packing and Packed Tower Design, 2nd ed., U.S. Stoneware Co.



NOMENCLATURE

PROPERTY	SYMBOL	BRITISH UNITS	METRIC UNITS	DEFINITIONS
Gas Rate	G	Lbs/ft² sec	KG/M² sec	$X = \frac{L}{G} \frac{\sqrt{\rho_G}}{\sqrt{\rho_L}}$
Liquid Rate	L	Lbs/ft² sec	KG/M² sec	
Gas Density	ρ_G	Lbs/ft³	KG/M³	$Cs = \frac{G}{\sqrt{\rho_G}\,\sqrt{\rho_L - \rho_G}} = \frac{Fs}{\sqrt{\rho_L - \rho_G}}$
Liquid Density	ρ_L	Lbs/ft³	KG/M³	
Liquid Viscosity	v	Centistokes	Centistokes	$Fs = \frac{G}{\sqrt{\rho_G}}$
Coefficient F	F	Packing Factor, dimensionless		

Figure 20.13E Latest version, generalized pressure drop correlation (GPDC). Used by permission of Norton Chemical Process Products.

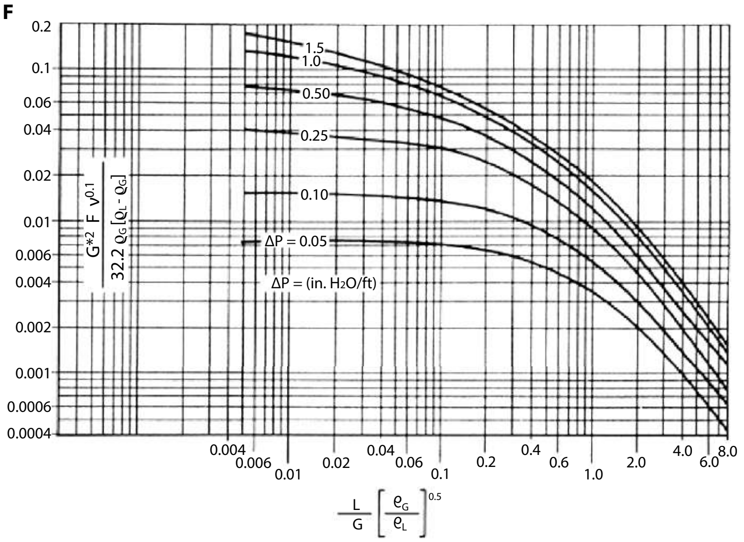


Figure 20.13F Strigle's latest generalized pressure drop correlation. Note G* = gas mass velocity, lb/ft²-s. Used by permission of Strigle, R. F., Jr., Packed Tower Design and Applications; Random and Structured Packings, 2nd ed. Gulf Publishing Co., © (1994), p. 19.

$$2.\ \text{Capacity parameter (CP), ordinate} = C_s\,F^{0.5}\,v^{0.5} \tag{20.16}$$

$$C_s = V_g\sqrt{\frac{\rho_G}{(\rho_L - \rho_G)}},\ \ \text{ft/sec., ft/s.} \tag{20.17}$$

where

C_s = capacity factor, ft/s
V_g = superficial gas velocity, corrected for densities ft/s
F = packing factor from Tables 20.25A–20.25G
L_h = liquid mass velocity, lb/ (ft^2) (h)
G_h = gas mass velocity, lb/(ft^2) (h)
ρ_G = gas mass density, lb/ft^3
ρ_L = liquid density, lb/ft^3
ν = kinematic liquid viscosity, centistokes = centipoise/(ρ_L/62.4)

Note differences in some symbol units for various GPDC charts.

For tower sizing:

1. Calculate using ordinate value = $C_SF^{0.5}\nu^{0.05}$
2. Calculate allowable gas mass velocity (ft/s), C_S

Flow parameter at maximum flow location:

$$FP = \frac{L}{G}\left[\frac{\rho_G}{\rho_L}\right]^{0.5} \tag{20.18}$$

Strigle [22] reports that the correlation of Figure 20.13F (and probably Figure 20.13G, because it is from the same data) predicts pressure drops to an accuracy of ±11 %, and suggests that this is probably the best attainable with available data. Better accuracy can be obtained only when using data specific to a particular packing family of sizes, such as the Nutter, Norton, and Glitsch respective packings noted hereafter. For Figures 20.13B through H, at high liquid rates, pressure drop may become somewhat greater than obtained from the charts (GPDC), particularly for the smaller packing sizes. For the higher liquid rates, the charts of the manufacturer of the packing should be used, rather than the GPDC. Strigle [22] reports that Kister and Gill's [23] tests indicate that from over 3,000 pressure drop measurements, the results fit Figure 20.13G for 80% (excellent) and another 15% (reasonable) fit.

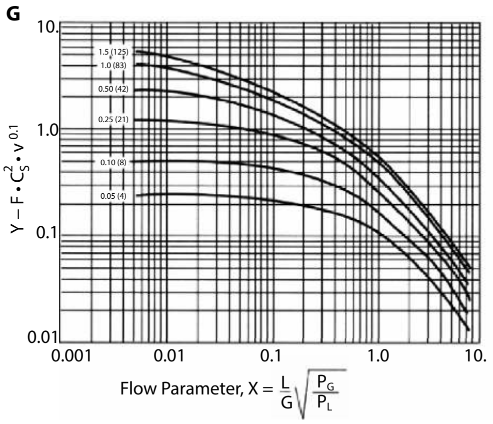
Strigle [2, 24] describes the hydraulics and HETP performance of a packed column by referring to Figure 20.14. As noted, the HETP values are essentially constant over a wide range of C_s values shown as B–C on the figure. Note that C_s can be expressed as

$$C_s = V_g\left[\rho_g/(\rho_l - \rho_g)\right]^{0.5} \tag{20.19}$$

or

$$= G/[\boldsymbol{\rho}_g\,(\boldsymbol{\rho}_l - \boldsymbol{\rho}_g)]^{0.5} \tag{20.20}$$

As vapor flow rate increases, so too does the contact between liquid and vapor. Hence, the rate of mass transfer will rise and the HETP value will improve. With increasing vapor rate, liquid entrainment into the vapor phase will occur and lower the efficiency (and raise the HETP) to the "maximum operating efficiency" [24] at point F where the C_s value rises above the efficiency used for design. Thus, the "maximum operating capacity" is well below any physical



Value for Coefficient F (Packing Factor)

IMTP Size	No. 15	No. 25	No. 40	No. 50	No. 70
F when C_S in m/s	549	441	258	194	129
F when C_S in ft/s	51	41	24	18	12

Symbols:
$C_s = V\sqrt{P_G/(P_L - P_G)}$, ft/sec, or m/sec.
V = Superficial gas velocity, ft/sec, or m/sec.
P_G= Gas or vapor density, lb/ft^3, or kg/m^3
P_L= Liquid density, lb/ft^3, or kg/m^3
L = Liquid mass rate, lb/hr, or kg/hr
σ = Surface tension, dyne/cm
μ = Liquid viscosity, centipoise
ν = Liquid kinematic viscosity, centistokes

Figure 20.13G Generalized pressure drop correlation for non-foaming systems for IMTP metal random packing. Parameter of curves is pressure drop in inches of water/foot packed height. Numbers in parentheses are mm of water/meter of packed height. Used by permission of Norton Chemical Process Products Corp. Bull-1HP-1, 12/91 (1987).

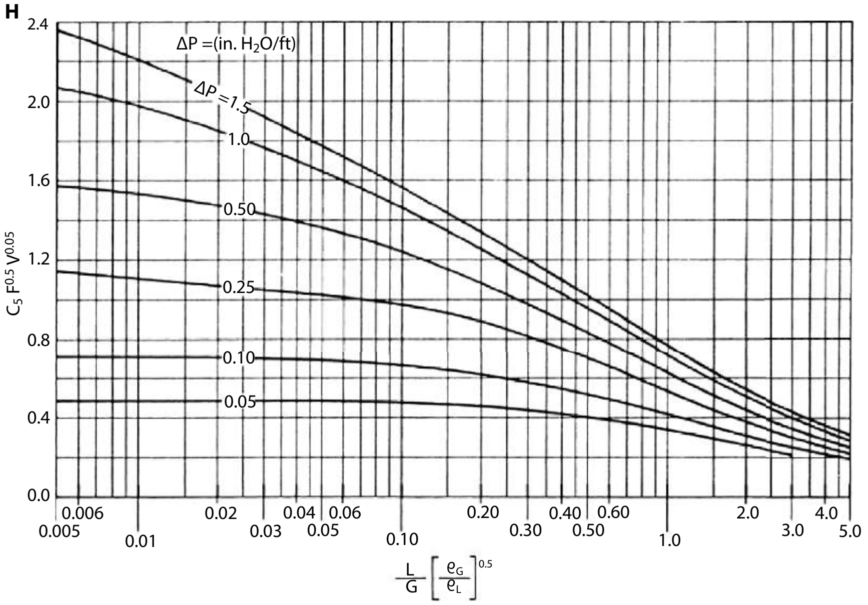
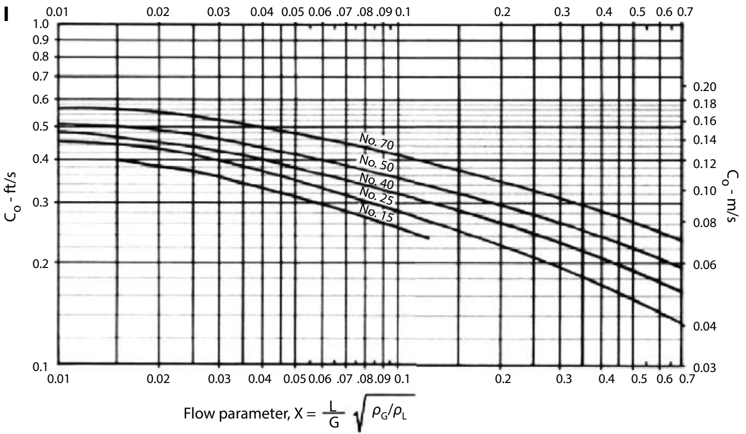


Figure 20.13H Updated generalized pressure drop correlation rearranged version of earlier Eckert and Leva, using linear scale for the ordinate and use of capacity factor, C_s. Used by permission of Strigle, R. F., Jr., Packed Tower Design and Applications; Random and Structured Packings, 2nd ed. © Gulf Publishing Co. p. 21 (1994). Note: G = gas, lb/ft^2 h, L = liquid, lb/ft^2 h.



L = liquid mass rate
G = gas mass rate
ρ_L = liquid density
ρ_G = gas density
V = superficial gas velocity, m/s or ft/s

Data Range

$5 \le \sigma \le 73$
$0.07 \le \mu \le 1.1$

$X = \frac{L}{G}\sqrt{\frac{\rho_G}{\rho_L}}$ Flow parameter

σ – surface tension, dyne/cm
μ – liquid viscosity, cp
ν – liquid kinematic viscosity, cs

Figure 20.13I Norton IMTP Packing, Efficient Capacity Correlation for random metal packing only for non-foaming systems. Norton recommends designing up to 90% of efficient capacity:

$$\text{Efficient Capacity, } C_{SC} = C_o\left[\frac{\sigma}{20}\right]^{0.16}\left[\frac{\mu}{0.2}\right]^{-0.11}, \ ft/s$$

Used by permission of Norton Chemical Process Products Corp., Bull. IHP-1 (9/87).

flooding point. In fact, the term "maximum operating capacity" is considered as a much more meaningful term to establish performance than "loading point" where earlier this was referred to as about point C [2]. The value of C_s at point D for atmospheric distillations has been found to occur at about 91% of the "maximum operating capacity" [24] at point F. The capacity factor C_s for design at point E has been set at the "maximum operating capacity," point F. The value of C_s for point E is approximately 87% of C_s at the maximum efficiency, point D. By setting the design capacity, C_s, as previously noted, the system should then be capable of operating up to 125% of design capacity and remain stable, and be conservative for mass transfer efficiency for vapor boil-up rates from point E to point F.

Thus, for mass transfer performance design, a specific design HETP value should be established, which in effect represents the range from point B through E, for C_s values above point F, the HETP values will be greater (and thus less efficient contact).

Some recent evaluations of data by other investigators indicate that a so-called "loading region" does not exist as clearly as may be suggested by other studies, and therefore, they suggest operations essentially up to the flooding point. For a good, reliable design that must allow for fluctuations in feed, and possibly column back-pressure upsets, designing to the flood region cannot be recommended. Design limits are discussed later.

Table 20.25A Packing factors,* ft^{-1} wet and dumped packing.

		Nominal packing size, in.										
Packing type	**Material**	**¼**	**3/8**	**½**	**5/8**	**¾**	**1 or #1**	**1 ¼**	**1 ½**	**2 or #2**	**3**	**3 ½ or #3**
Intalox® IMTP®	Metal				51 (#15)		41 (#25)		24(#40)	18 (#50)		12 (#70)
Hy-Pak™	Metal						45		29	26		16
Super Intalox® Saddles	Ceramic						60			30		
Super Intalox Saddles	Plastic						40			28		18
Pall Rings	Plastic				95–102*		52*–55		31*–40	26	18*	17
Pall Rings	Metal		76*–81		76*		48*–56		33*–40	23*–27		18
Intalox® Saddles	Ceramic	725	330	200		145	92		52	40		22
Raschig Rings	Ceramic	1600	1000	580	380	255	179	125	93	65		37
Raschig Rings	1/32″	700	390	300	170	155	115					
	Metal											
Raschig Rings	1/16″											
	Metal			410	300	220	144	110	83	57		32
Bed Saddles	Ceramic	259		240		170	110		65	45		
Snowflake™	Plastic							13 (#38)		13 (#50)		13 (#90)

Packing factors determined with an air–water system in 30″ I.D. Tower.

Updated by permission from R.F. Strigle, Jr., *Random Packings and Packed Towers,* Gulf Publishing Co. (1987), added Snowflake™ data by permission

*By permission Norton Co., from data compiled in Norton Co. Laboratories, Copyright 1977.

Table 20.25B Koch packing factors*.

Packing type	Material	Nominal packing size, in										
		¼	**3/8**	**½**	**5/8**	**¾**	**1 or #1**	**1 ¼**	**1 ½**	**2 or #2**	**3**	**3 ½ or #3**
Flexisaddles	Plastic						30			20		15
Flexirings	Plastic[†]				78		45		28	22		18
Flexisaddles	Ceramic	600		200		145	98		52	40	22	

*By permission, Koch Engineering Co. Inc.

[†] Use for plastic or metal.

Table 20.25C Glitsch packing factors.

Packing type	Material	Nominal packing size, in.										
		¼	**3/8**	**½**	**5/8**	**¾**	**1 or #1**	**1 ¼**	**1 ½**	**2 or #2**	**3**	**3 ½ or #3**
Ballast™ Ring	Metal						48		28	20		15
Ballast™ Ring	Plastic				97		52		32	25		16
Ballast™ Saddle	Plastic						30			20		15

*By permission, Glitch Inc.

Table 20.25D Glitsch packing factors for cascade® mini-rings.

Size	Plastic	Ceramic	Metal
0	–	–	55
0A	60	–	–
1	29	–	40
1A	30	–	–
1.5	–	–	29
2C	15	38	22
2A	30	–	–
2B	18	–	–
2C	19	–	–
2.5	–	–	19
3	11	24	14
3A	12	–	–
4	–	–	10
5	–	18	8
7	–	15	–

Used by permission: Glitch, Inc., Bull. 345.

Table 20.25E* Nutter Rings™ random packing.

Size no.	Pieces/ft³	Ft²/ft³	Lb/ft³	%void	Packing factor**	Relative HETP values
0.7	4,740	69	11.0	97.8	xx	0.72
1.0	1,900	51	11.1	97.8	30	0.83
1.5	760	38	11.3	97.8	24	0.94
2.0	385	29	10.8	97.9	18	1.00
2.5	250	25	9.0	98.2	xx	1.18
3.0	120	20	8.3	98.4	xx	1.40

*Nutter use their own proprietary computer program and not the conventional GPDC Chart shown in Figures 20.60-69.
**Values shown developed by Kister [90] from data supplied by Nutter Engineering Co., a Harsco Corp.
Used by permission of Nutter Engineering Co., a Harsco Corp.; Bull. NR-2.

Table 20.25F Packing factors (PF) dumped packing.

Packing Type	Material +	Nominal Packing Size (inches)										
		¼	3/8	½	5/8	¾	1	1.25	1.5	2	3	3.5
IMTP®	Metal				51		40		24	18	12	
Hy-Pac™	Metal						45		29	26		16
Super-Intalox Saddles®	Ceramic						60			30		
Super-Intalox Saddles®	Plastic						40			28		18
Pall Rings	Plastic				75		55		40	26		17
Pall Rings	Metal				70		56		40	27		18
Intalox Saddles®	Ceramic	725	330	200		145	92		52	40	22	
Raschig Rings	Ceramic	1600	1000	580	380	255	155	125	95	65	37	
Raschig Rings	1/32″ Metal	700	390	300	170	155	115		95			
Raschig Rings	1/16″Metal			410	300	220	144	110		62	32	
Berl Saddles	Ceramic	900		240		170	110		93	45		
Flexiring	Metal						49		65	23		16
Fleximax	Metal						35		26	17	11	
Cascade Mini Rings	Metal						39	36	33	26	18	
Cascade Mini Rings	Plastic						44		33	20	17	

Note: Values in this table are average values for the packing factor (PF). PF is actually a function of loading. Specific correlations for each packing from the vendors.
Source: Gas Processors Suppliers Association (GPSA), 12th. Edition 2004, GPSA, Tulsa, Oklahoma, Used by permission of GPSA, All rights reserved.

Table 20.25G Packing Factor Values (1/ft.) for Structured Packings.

	Flexipac		Gempak		Sulzer		Munters		Intalox	
	2	4	2A	4A	CY	BX	1260	19060	2T	3T
F	22.6	6	16	32	70	21	27	15	17	13

Source: Wankat, P.C., Separation Process Engineering 2nd Ed., Prentice Hall, 2007.

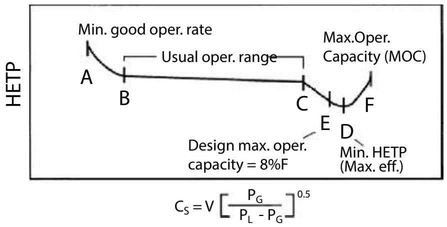


Figure 20.14 Typical HETP curve illustrating operating and design relationships. Normally reasonably constant over wide range of vapor flows. Note: C_S = capacity factor; V = gas mass velocity-lb/ft^2 s. Adapted by permission from Strigle, R. E. Jr., and Rukovena, F., and reproduced with permission of the American Institute of Chemical Engineers. Chemical Engineering Progress, Ma. (1979). All rights reserved.

20.13 Flooding Point

At the second sharp change in the slope of the pressure drop curve, Figure 20.12, the packing tends to hold up more and more liquid as the gas flow increases. This creates a rapid increase in pressure drop. The flooding point of the system is said to be the point of the second inflection of the pressure drop curve. Here, the liquid build-up on top of the packing increases and the pressure drop essentially becomes infinite for a finite increase in gas rate. In many actual cases, operation can be maintained at the flooding point, but it will be erratic, the performance (efficiency) of operation poor, and the entrainment carry-over excessively high. It is obvious that towers are not designed for flood point operations, but at 40–60% of gas and liquid rates associated with this point. Figure 20.13D indicates that the flooding region usually is above 2.5 in. water/ft, but note how cramped and much more sensitive this condition becomes as the extreme right-hand side of the graph is approached.

Kister's [23] study indicates that flooding with the newer packings occurs just below 2 in. water/ft of packing, somewhat below the prediction given by the earlier Eckert chart, Figure 20.13C or D. The data plotted by Kister indicate that for larger packings (2 and 3 in., for example), the flooding point is found to be at much lower pressure drops than suggested by Figure 20.13C. The pressure drop at flood point has been found to be independent of the Flow Parameter on the charts, but does vary with the packing family and packing types [23, 25].

Kister [23] has developed a new approach at establishing the flood point that appears to suit the available data and is significantly more accurate than reading the upper curve on Figure 20.13C.

$$\Delta P_{\text{flood}} = 0.115 F_p^{0.7} \,(\text{do not extrapolate below } F_p = 14) \tag{20.21}$$

where

ΔP_{flood} = pressure drop at flood point for all random packings, in. water/ft of packing
F_p = packing factor, empirical, based on packing size and shape. 1/ft, see Tables 20.25A–F.

By calculating ΔP_{flood}, the capacity parameter can be determined using the calculated flow parameter and Figure 20.13E, and, if available, the SLE (Sherwood, Leva, Eckert) charts of Reference 23. For $F_p > 60$, the calculated ΔP_{flood} using the equation coincides with Eckert's Figure 20.13C flood line. Figure 20.15 illustrates the relationship between packing factors and flooding pressure drop, and is represented by the ΔP_{flood} equation. The predicted results of the equation are ±10 to 15% based of the plotted data [23].

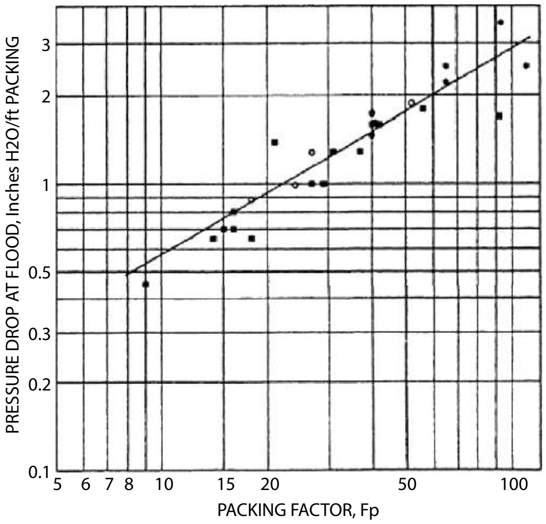


Figure 20.15 Flood pressure drop vs. packing factor for random packings. Reproduced with permission of the American Institute of Chemical Engineers, Kister, H. Z., and Gill, D. R., Chemical Engineering Progress, V. 87, No. 2 © (1991); all rights reserved.

20.14 Foaming Liquid Systems

For an accurate design, the effects of foaming of the liquid as it flows through the random packed/structured bed should be known, estimated, or determined by experimentation. There is little published data on the subject except Eckert [26] and Strigle [2]. Generally, foaming systems produce higher pressure drops than non-foaming, most probably due to the blocking of packing voids by the foam. Therefore, it is wise to determine the foaming behavior of the specific system, and due to the lack of published data, generous allowances for a pressure drop should be made in design, perhaps even two to three times greater than expected.

Hsu [27] presents equations for directly calculating random packings based on published data and which are adaptable for computer programming. These are very useful for studying the effects of variables. The basic data are essentially a match with Figure 20.13D.

20.15 Surface Tension Effects

Strigle [2] reports that there is no broadly documented agreement of the surface tension effects on the capacity of packed beds. Eckert [2, 26] concluded that surface tension of a non-foaming liquid had no effect on capacity. These results were later confirmed. For absorption systems, these results also hold [2].

For hydrocarbons in high-pressure fractionators, Strigle [2] reports there is aeration of the rather low surface tension liquid phase. This effect increases as surface tension increases and as the vapor density increases. This reduces the effective liquid density, and so the volume occupied by a given mass of liquid is increased. This aeration effect can vary from 0.9 at atmospheric pressure for non-foaming liquids to 0.7 for hydrocarbon systems (not absorbers operating at 35% of critical pressure).

Kaiser [28] presents a correlation analysis for flooding in packed towers based on single phase hydraulics. However, Kaiser's article does not present a comparison of results with the other more conventional techniques.

20.16 Packing Factors

The use of "packing factors" is well established in the design concepts of evaluating packed tower performance. Essentially all of the manufacturer's published data are for "wet and dumped" packing factors, F. Robbins proposes using only "dry" factors. This factor is a unique characteristic of each packing size and style/design. These factors cannot be determined by calculation from the physical dimensions; they are determined experimentally.

Packing factor selection significantly affects the performance of a packed tower system. These factors are only suitable for discreet particle type packing, and their values vary with the way the packing is installed in the tower. For example, the factors for a ceramic packing are different in the case of packing floated (dumped) into a tower full of water and the particles allowed to float down, compared to the same packing dumped into a dry empty tower (where significant breakage can occur and consolidate the packing), or even to packing "hand-placed" or stacked dry.

Often it is only necessary to change a packing size or type to modify the capacity and/or contacting efficiency of an existing tower, because this change affects the packing factor. Tables 20.25A–G present specific packing factors from the manufacturers.

Many of the packings of the various manufacturers are essentially identical in shape, size, and performance factors. Some packing manufacturers suggest adjusting packing factors for vacuum and pressure distillations; however, this should only be done after consultation.

The experimentally determined packing factors are the only reliable values to use for design calculations, although estimates can be made for packing shapes when no data are available. The packing characteristic is expressed as

$$F = a/\varepsilon^3 \qquad (20.22)$$

where

a_t = specific surface of packing, ft^2/ft^3
a = effective interfacial area for contacting, ft^2/ft^3
ε = fractional voids

The values of a/ε^3 determined experimentally by Lobo *et al.* are indicated [29]. These are the values in the development of the basic relation expressed in Figure 20.13A with correction of ψ^2 suggested by Leva [30]. These a/ε^3 values were found to correlate a considerable amount of the literature data within 12%. This would mean about a 6% error in tower diameter determined at flooding conditions.

Lobo *et al.* [29] proposed the packing factor, F, and experimentally determined that it better represented the data than the calculated a/ε^3 term. Values calculated using surface area per cubic foot and percent free gas space from manufacturer's tables can be in error by as much as 40%. The values are dependent upon the method of packing the tower, i.e., dry dumped, wet dumped, or wet dumped and shaken. The latter condition may approximate the situation after a tower has been running a while and the packing settled.

Experience definitely indicates that the packing factor, F, increases with hours of operation for ceramic materials up to some limit. This is due to settling, breakage, plugging, etc. For design of commercial towers, values of F should be increased from 15 to 75% for ceramic materials, over values read from Tables 20.25A–E. The percent increase depends upon the tendency of the shape to disintegrate into smaller pieces during operations-flooding, gas surging, etc. In general, circular shapes exhibit the least tendency to break up. As a reasonable value where data are available, the average of the wet dumped, and wet-dumped-and-shaken values for tower voidages are recommended.

Leva [31] has converted the data of Lubin into correction factors to apply to a non-irrigated bed pressure drop in order to derive a pressure drop for a liquid–gas system in the loading to flooding range. In general, this does not appear any more convenient to use than Figure 20.13D.

Relations expressing the fractional voids in a ring packed bed are useful in estimating the "ε" values for a/ε^3 determinations [29]. The average deviation is ± 2.6%.

Dry packed tower:

$$\varepsilon = 1.046 - 0.658\,\phi \qquad (20.23)$$

Wet packed, unshaken tower:

$$\varepsilon = 1.029 - 0.591\,\phi \qquad (20.24)$$

Wet packed and shaken tower:

$$\varepsilon = 1.009 - 0.626\,\phi \tag{20.25}$$

where

$$\phi = \frac{1-(d_i/d_o)^2}{\left(l d_o^2\right)^{0.017}}, \quad \text{not valid if } \phi < 0.20 \text{ or for extra thick walls or solids}$$

l = ring height, in.
d_o = outside diameter of ring, in
d_i = inside diameter of ring, in.

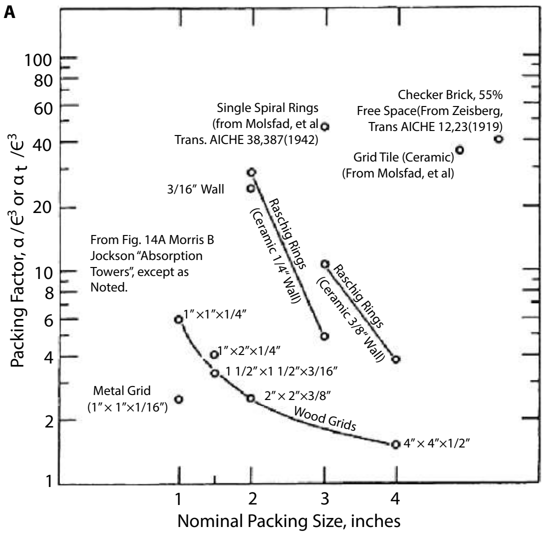


Figure 20.16A Packing factors (stacked packing selected grids). Used by permission of Morris and Jackson, Absorption Towers Butterworth Scientific Publications, and Imperial Chemical Industries, Ltd., and adapted by U.S. Stoneware Co. (now Norton Chemical Process Products Corp.).

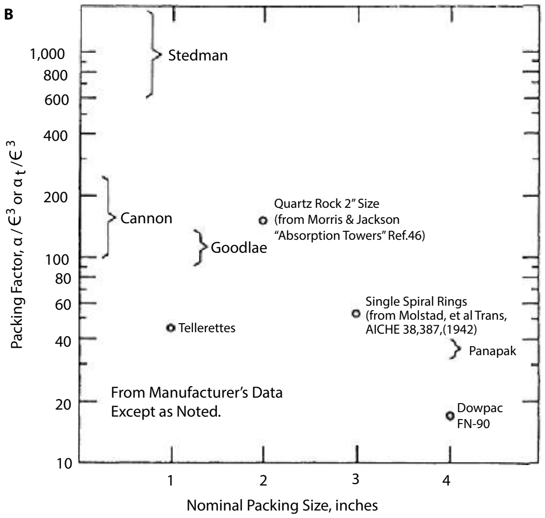


Figure 20.16B Packing factors (screen packing and random dumped packing). Used by permission of U.S. Stoneware Co. (now Norton Chemical Process Products Corp.).

The generalized correlations of Sakiadis and Johnson [32] are reported to satisfy a wide variety of systems.

Manufacturers of commercial packings provide packing factors for their products. Many (not necessarily all) of the commonly used packing factors are presented in Table 20.25A. These values are to be used with the application of Figures 20.13B and 20.13F. Factors presented in Figures 20.16A and 20.16B can also be used where the design requires them.

20.17 Recommended Design Capacity and Pressure Drop

The relationships in packed tower performance, which are concerned specifically with the gas and liquid flows through a bed, are expressed as a function of pressure drop. Pressure drop may be created by poor packing arrangements, i.e., tight and open sections in the bed, breakage of packing and settling of the bed, or plugging of void spaces by solids or reaction products. All of these are in addition to the inherent characteristic resistance of a particular packing to flow of fluids. This resistance will be different if the system is single-phase rather than the two-phase systems existing in most distillation, adsorption, scrubbing, or desorption operations. The basic pressure drop performance pattern of nearly all packings follows the same form as Figure 20.12.

Below the loading region, the pressure drop can be read from appropriate system curves if available, as Figure 20.12. However, for general use, the data have been well correlated (Figures 20.13B–F). The slope of most of the

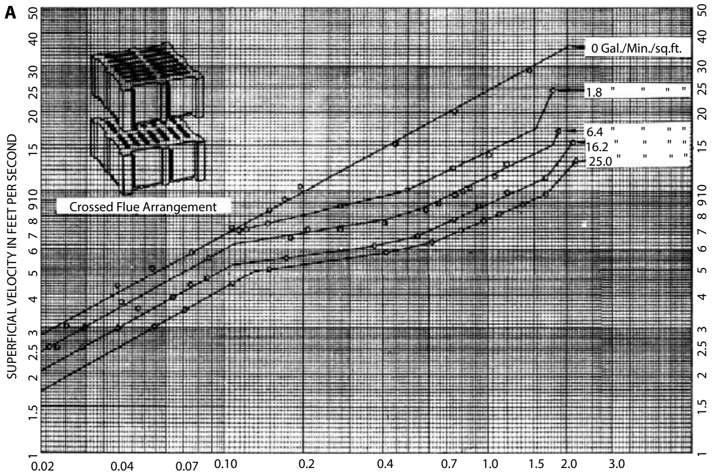


Figure 20.17A Pressure drop in inches of water per foot of height, drip point tile shape 6295 with crossed flue arrangement. Used by permission of General Refractories Co.

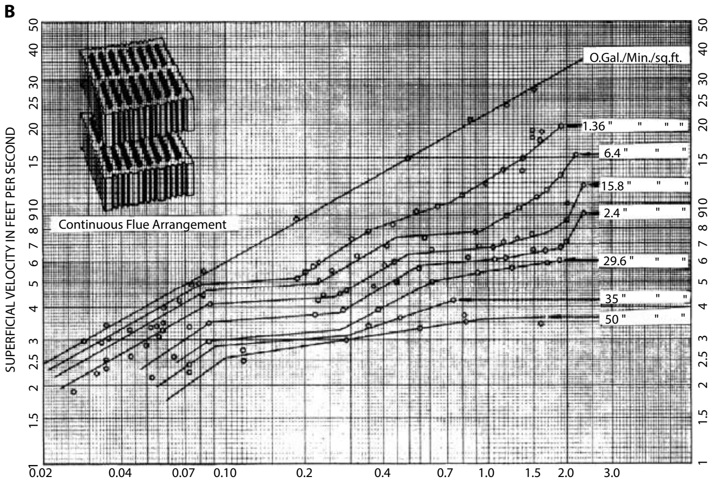


Figure 20.17B Pressure drop in inches of water per foot of height, drip point tile shape 6897 with continuous flue arrangement. Used by permission of General Refractories Co.

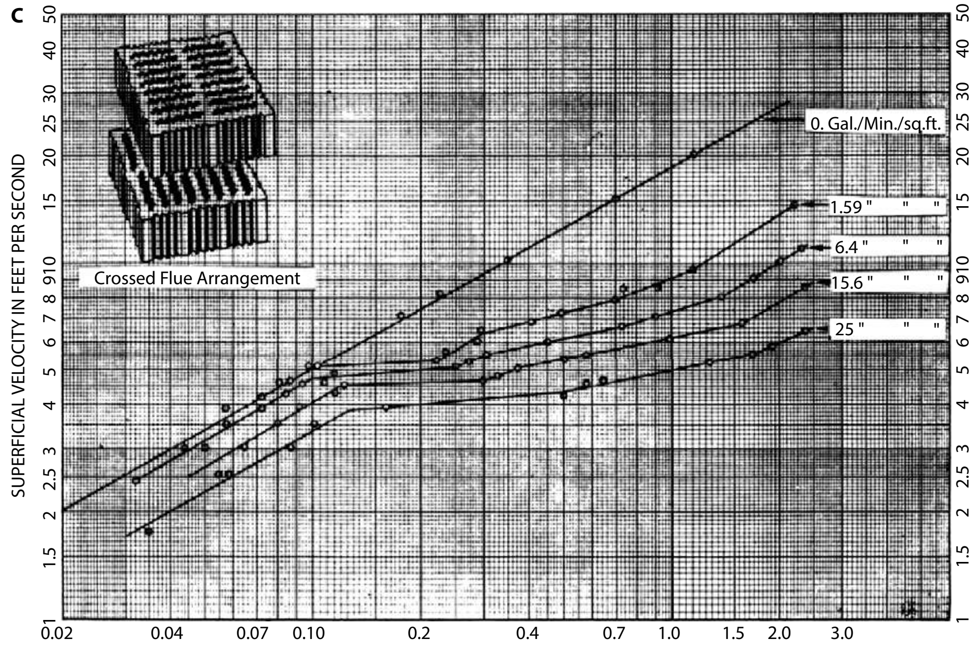


Figure 20.17C Pressure drop in inches of water per foot of height, drip point tile shape 6897 with crossed flue arrangement. Used by permission of General Refractories Co.

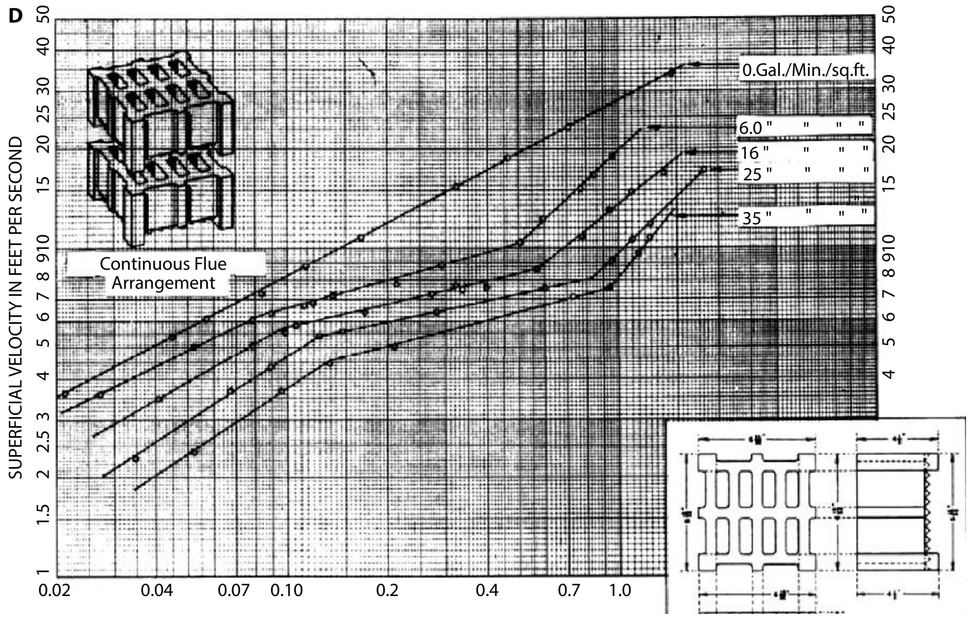


Figure 20.17D Pressure drop in inches of water per foot of height; drop point tile shape 6146 with continuous flue arrangement. Used by permission: General Refractories Co.

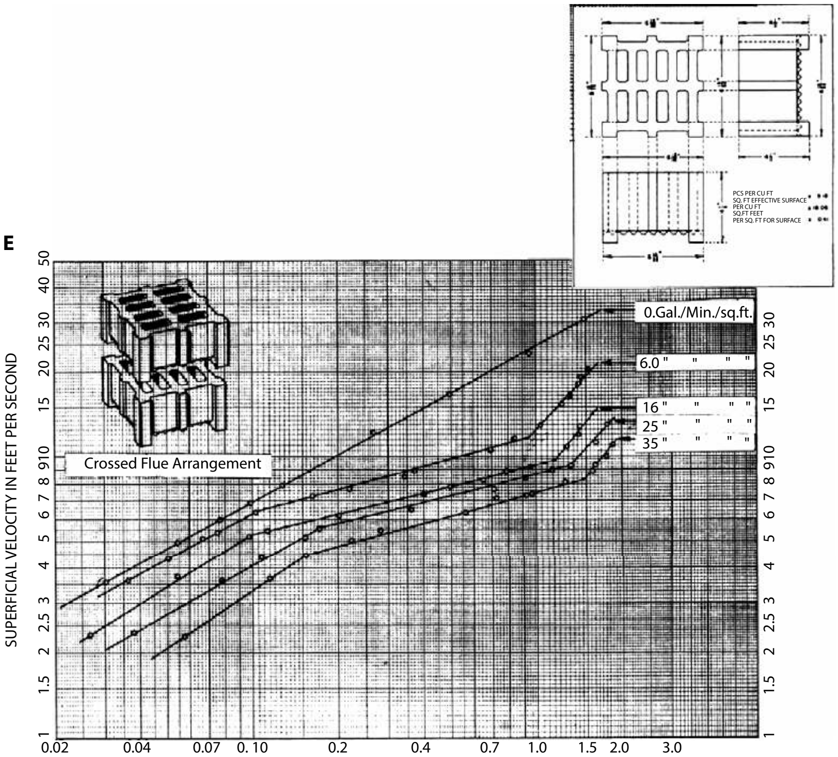


Figure 20.17E Pressure drop in inches of water per foot of height, drip point tile shape 6146 with cross flue arrangement. Used by permission of General Refractories Co.

pressure drop curves indicates a proportionality of 1.8 to 2.8 power of the superficial gas mass velocity up the tower. This performance is typical of the gas-continuous range of packed tower operation. In Figure 20.12, the curves for performance at water rates of L' = 1,250 through L' = 9,950 all seem to be a part of the same family. The curve L' = 12,500 will be discussed later. The pressure drop information to follow is valid only for the gas-continuous type of operation. Fortunately, the majority of packed towers operate in this condition; however, the liquid-continuous will be considered later.

Pressure drop data for several styles and arrangements of drip point grid tile are given in Figures 20.17A–E. These are not included in the general GPDC correlations for random packings.

Figures 20.13E, F, and H are about the latest general-purpose correlations presented by several manufacturers of packing materials. The relative differences between the various correlations appear to be minor, thereby allowing any packing performance to be evaluated on any chart, as long as the packing factors, F, have been determined on the same basis. Packing factors are presented in Tables 20.25A and B and relate to discreet particle packings similar to those illustrated in Figures 20.6A–X. The compacted, structured, or grid packing materials such as shown in Figures 20.6Y to 20.6UU do not use the same packing factor concept for design evaluation. Therefore, the respective manufacturer should be consulted when using proprietary materials such as those in Figures 20.6Y to 20.6UU since they have their own design/rating technique.

20.18 Pressure Drop Design Criteria and Guide: Random Packings Only

A. Determine the height of packing required (not a function of diameter) from HETP in distillation section and Kga in absorption section. This provides the total height of packing required, considering the packing efficiencies and minimum wetting requirements, previously discussed. From this total height, the expected total column pressure drop can be established, recognizing the pressure loss through support plates, distribution devices, etc.

B. Calculate the abscissa of Figure 20.13F; for example, $L/G\sqrt{\rho_G/\rho_L}$ or other abscissa values.

C. Select a design/operating pressure drop, as shown on the curves of Figure 20.13F. Suggested selection basis is as follows:

1. Low to medium pressure column operation, select design pressure drop of 0.40 to 0.60 in. water/ft of packing height, although some towers will operate at 1.5 to 2.0 in. water/ft. Select a design C_s above point C and below point D (Figure 20.14).
2. Absorption and similar systems; select pressure drop of 0.1 to 0.4 in. water/ft of packing, or 0.25 to 0.4 for non-foaming systems. For "median" foaming systems, use a maximum of 0.25 in. water/ft for the highest loading rate [2]; see Table 20.26. The values in the table should be reduced for high viscosity fluids. At rates above 20, limit the gas rate to 85% of the rate, which would give a pressure drop of 1.5 in. water/ft referring to the GPDC (Figure 20.13F).
3. Atmospheric or pressure distillation, select pressure drop of 0.50–1.0 in. water/ft.
4. Vacuum distillation varies with the system and particularly with the absolute pressure required at the bottom of the column; normally select low pressure drop in the range of 0.1 to 0.2 in. water/ft of packing. For in vacuum service of 75 mm Hg and lower, the pressure drop obtained from the GPDC (Figures 20.13F and 20.13G; for Norton's IMTP packing only) can be up to 20% higher than industrial experience for the same C_s values at a flow parameter (FP) of 0.01 [2]. At an absolute column pressure of 10 mm Hg or less, the pressure drop actually can be up to 30% lower than that read from Figure 20.13F at the FP of 0.01 and the same C_s value [2]. The conclusion is that the generalized correlation (Figures 20.13F and 20.13G) always gives a conservative design ΔP at operating pressures less than 70 mm Hg abs. Several other factors must be considered, such as variability of gas and liquid rates and densities or specific volumes.
5. Foaming materials should be operated at 0.1–0.25 in. water/ft, or better, obtain data. Actual ΔP/ft may run two to five times chart values.
6. Vacuum service requires recognition of minimum liquid flow; refer to section on minimum wetting rate. Pressure drops are designed to be low, but normally not lower than 0.10 in. water/ft.
7. Stripping gas from a liquid phase; pressure drop is usually set to be 0.15 to 50 in. water/ft packing at the flow rates for maximum loading or maximum operating capacity. For foaming systems, set pressure drop at ≤ 0.25 in. water/ft using inert gas. For steam use, set pressure drop ≤ 0.3 maximum in. water/ft for medium foaming conditions [2].
8. For high-pressure distillation of light hydrocarbons, industrial performance indicates that the pressure drop actually obtained is about twice that predicted by the use of the GPDC charts (Figures 20.13F and 20.13G; for Norton's IMTP packing only). When the vapor density is at least 6% of the liquid density, the actual pressure drop is expressed as [2]

$$\Delta P = \frac{33F^{0.5}(C_s)^{2.4}}{\sigma} \text{ in. water/ft packing} \tag{20.26}$$

Table 20.26 Maximum recommended liquid loading.

Packing size (in.)	Liquid rate (gpm/ft^2)
¾	25
1	40
1 ½	55
2	70
3 ½	125

Used by permission of Strigle, R. F., Jr., "Random Packing and Packed Towers" © (1987), Gulf Publishing Co., all rights reserved.

where

C_s = capacity factor, ft/s
σ = surface tension, dynes/cm
F = packing factor

For non-hydrocarbon systems, Strigle [2] recommends the general outline to follow, but more detail for a specific design may be obtained from the reference as well as the manufacturer:

1. Select a design C_s vapor rate as a percent of the MOC (maximum operating capacity) of the packing.
2. Adjust maximum C_s required, adjusting for:
   a. Effect of liquid viscosity: The maximum operating capacity varies as the 0.2 to 0.25 power of surface tension of the liquid phase.
   b. Packing capacity increases as the reciprocal of liquid viscosity to the 0.1 to 0.13 power. This effect is limited to liquid viscosities not < 0.09 cPs.
   c. Select design C_s at 80 to 87% of the MOC, C_s for the packing.
   d. When the gas density exceeds 6% of the liquid density, the pressure drop should be examined by Equation 20.26.
   Pressure drop should not exceed (at maximum operating capacity) [22, 24]

$$\Delta P_{max} /> (0.019)\ (F)^{0.7}, \text{ in. water/ft packing} \tag{20.27}$$

   or, from Kister and Gill [22, 23], pressure drop at flood for random dumped packing:

$$\Delta P_{flood} = (0.115)\ (F)^{0.7}, \text{ in. water/ft packing} \tag{20.21}$$

   e. For pressures approaching the critical, contact the packing manufacturer for performance guidance.
   f. Select a design C_s that provides allowances for operational and capacity fluctuation and/or surges; 15–25% above the design rate is usually adequate [2], making the design C_s from 80 to 87% of the maximum operational C_s for the specific packing. The maximum operational capacity, MOC, $C_{s,max.}$ is the point, C (Figure 20.14), where the vapor rate has increased to a point where it begins to interact with the liquid. In previous publications, this point was termed "loading point." Actually, for rates slightly greater than point C, the efficiency of the packing increases (HETP becomes smaller).
3. Select a packing and determine its packing factor from Tables 20.25A–E. Packing is selected for its expected process HETP or Kga performance, pressure drop, and materials of construction for the system. Table 20.17 presents summary comments for applications. Selection guides are as follows:
   a. As packing factor, F, becomes larger by selection of smaller sized packing; gas capacity for the column is reduced; and pressure drop will increase for a fixed gas flow.
   b. Some packings are sized by general dimensions in inches, while some shapes are identified by numbers, #1, #2, #3 for increasing size.
   c. Not all packings are manufactured in all materials of construction, i.e., ceramic, various plastics, various metals.
   d. Packing size versus tower diameter recommendations; general guides not mandatory, base selection on performance.

Tower Diameter, ft	**Nominal Packing Size, in.**
<1.0	<1
1.0–3.0	1–1 1/2
>3.0	2–3

Table 20.26 shows what Strigle [2] recommends for the maximum liquid loading as related to packing size.

a. Referring to Figure 20.13B or 20.13C, read up from the abscissa to the pressure drop line selected, and read across to the ordinate (note differences):

$$\text{Ordinate No.} = \frac{G^2 F \mu^{0.1}}{\rho_G(\rho_L - \rho_G)g_c}, \text{(Figure 20.13C)} \tag{20.28}$$

or

$$= \frac{G^2 F \nu^{0.1}}{(\rho_G[\rho_L - \rho_G])}, \text{(Figure 20.13B)} \tag{20.29}$$

or

$$= C_s\, F^{0.5}\, \upsilon^{0.05}, \text{(Figure 20.13H)} \tag{20.30}$$

Note units change for Figure 20.13G, where

$$C_s = V_g[\rho_g/(\rho_L - \rho_G)]0.5, \text{ft/s} \tag{20.31}$$

υ = kinematic liquid viscosity, centistokes

Substitute F and the other knowns into the equation and solve for G, the gas mass flow rate, lbs/ft^2 s, or G_g, lb/h-ft^2, as applicable.

Then, determine the required tower cross-section area and diameter:

$$\text{Diameter, ft} = 1.1283\left[\frac{\text{Gas rate, lb/sec}, G''}{G, \text{lb/sec}-\text{ft}^2}\right]^{1/2} \tag{20.32}$$

20.19 Effects of Physical Properties

For nonfoaming liquids, capacity of a packing is independent of surface tension. Foaming conditions reduce capacity significantly, and design should recognize this by selecting an operating pressure drop at only 50% of that for a normal non-foaming liquid.

For liquids of viscosity of 30 centipoise and lower, the effect on capacity is small. For high viscosity liquids, select larger packing to reduce pressure drop, and also consult the packing manufacturer.

Robbins' [33] correlation for pressure drop in random particle packed towers is based on a "dry packing factor," F_{pd}, whereas most of manufacturer's published values derived F (packing factor) from wet and dumped data, such as is used in Figures 20.13A-F and 20.13H. Reference to the tables in Robbins' presentation indicates that the differences between the dry, F_{pd}, and the wet and dumped, F (such as Table 20.25A), may be small, being from 0 to 10 points, averaging about 2–3 points lower. The packing manufacturer should be consulted over what dry packing factors to use in Robbins' method. The "dry" data simply means that there is no liquid (but gas) flowing. Robbins [33] lists values of F_{pd} for metal, plastic, and ceramic packings.

Dry bed pressure drop [33]:

$$\Delta P = C_o \rho_g V_s^2 = C_o F_s^2 = C_o G^2/\rho_g \tag{20.33}$$

Values of C_o come from Leva [31].

Robbins' new equation for generalized pressure drop for random tower packings:

$$\Delta P = C_3 G_f^2 (10^{2.7\times10-5[L_f]}) + 0.4\left[L_f/20{,}000\right]^{0.01} \times \left[C_3 G_f^2 \left(10^{2.7\times10-5[L_f]}\right)\right]^4 \qquad (20.34)$$

$$G_f = G[0.075/\rho_g]^{0.5}[F_{pd}/20]^{0.5} = 986F_s[F_{pd}/20]^{0.5} \qquad (20.35)$$

$$L_f = L\left[62.4/\rho_L\right]\left[F_{pd}/20\right]^{0.5} \mu^{0.1} \qquad (20.36)$$

The method as described by Robbins [33]:

1. For operating pressures above 1 atm, multiply G_f (Equation 20.35) by $(10^{0.3})(\rho_g)$.
2. For small packings with F_{pd} of 200 or greater, substitute $\mu^{0.2}$ in previous equation for $\mu^{0.1}$.
3. For large packings with F_{pd} below 15, use $(20/F_{pd})^{0.5}$ in place of $[F_{pd}/20]^{0.5}$ in previous equation for L_f, Equation 20.36.
4. Dry bed pressure drop:

$$F_{pd} = 278(\Delta P_{db})/F_s^2 \qquad (20.37)$$

Dry bed pressure drop values usually run 0.1 to 0.5 in. water/ft of packing [33]. Use Equation 20.34 when L_f is below 20,000. Packings operate essentially dry when L_f is below 1,500 (about 3 gpm/ft^2) at F_p = 20. Pressure drop at flooding is suggested to be predicted by Kister and Gill's relationship [23] presented in this text.

Robbins [33] suggested that random packed column design is similar to others presented in this text, but highlighted to determine the diameter of packed column:

1. Establish liquid and vapor rates.
2. Determine fluids physical properties.
3. Select design pressure drop for operations. Suggested values of below 1.0 in. H_2O/ft. Low-pressure, atmospheric, and pressure columns usually require 0.5 to 0.7 in. H_2O/ft, with absorbers and strippers around 0.2–0.6 in. H_2O/ft. For vacuum distillation, low values of 0.05–0.6 in. H_2O/ft are often necessary, usually depending on the required boiling point of the bottoms.
4. Calculate L_f/G_f; values of F_{pd} cancel out.
5. Using Equation 20.34 and trial and error, calculate using G_f, for given L_f/G_f until the desired ΔP is obtained. L_f must be below 20,000. For higher L_f, use chart in the original article (not included here).
6. Select packing for column and establish packing factor, F_{pd}.
7. Calculate column cross-section area using the operational gas rate, G, and the calculated value of G_f (gas loading factor);

$$G = G_f/\left[\left(0.075/\rho_g\right)^{0.5}\left(F_{pd}/20\right)^{0.5}\right], \text{lb/hr/ft}^2 \qquad (20.38)$$

8. Establish tower diameter; Robbins [33] recommends that the tower diameter should be at least eight times the packing size; if not, repeat the calculations with different packing.

where

A = tower cross-sectional area, ft^2
G_o = constant specific to a particular packing
G_1 = constant specific to a particular packing
G_3 = 7.4 x 10^{-8}
G_4 = 2.7 x 10^{-5}
D = tower diameter, ft

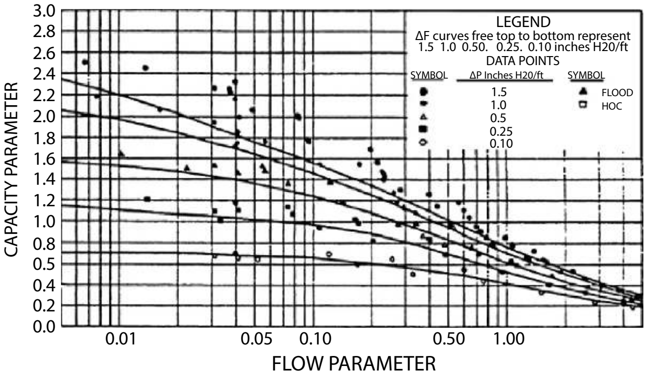
F_p = packing factor, dimensionless
F_{pd} = dry-bed packing factor, dimensionless
F_s = $V_s (\rho_g)^{0.5}$, (ft/s) $(\text{lb/ft}^3)^{0.5}$
G = gas loading, lb/h-ft^2
G_A = design vapor flow rate, lb/h
G_f = gas loading factor
L = liquid loading, lb/h-ft^2
L_f = liquid loading factor
ΔP = specific pressure drop, in. water/ft of packing
ΔP_{pb} = specific pressure drop through dry bed, in. H_2O/ft of packing
V_s = superficial gas velocity, ft/s.
ρ_g = gas density, lb/ft^3
ρ_L = liquid density, lb/ft^3
μ = liquid viscosity, cP.

Strigle [2] and Kister [23] point out the importance of evaluating data (where available) to reduce the need for interpolating the GPDC charts. The question of reasonably accurate (±10 to 15%) flooding pressure drop data has been studied by Kister [23], and the results suggest that the establishment of flooding pressure drop curves such as in Figure 20.13C may not be consistently accurate due to the variations in data used for correlation, i.e., data just as flooding begins, and then at full flooding. The data presented are for gas–liquid systems and not liquid–liquid extraction as illustrated by Strigle [2].

Strigle [2] identifies a regime 20% above point F on Figure 20.14 as the maximum hydraulic capacity and is termed the flooding point for atmospheric operations.

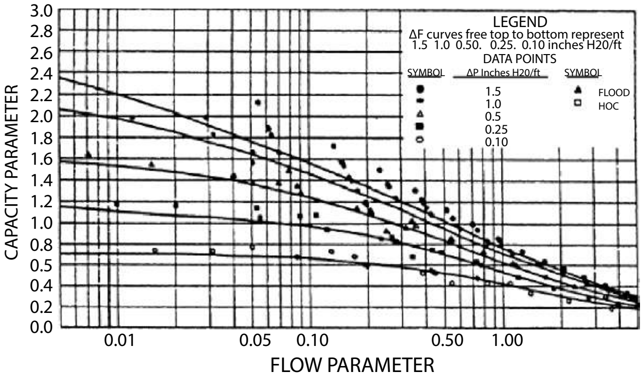
Kister [23] has correlated large quantities of available data for flooding and offers a new correlation based on the same GPDC correlation, i.e., the Sherwood–Leva–Eckert (SLE) correlation chart as developed by Strigle [2] (Figure 20.13H) for semi-log plot. Kister [19, 23] has presented plots showing the plotted data points on the Strigle or SLE charts for a wide selection of packing. In effect, this illustrates how the data fit the generalized charts.

As an example, Figures 20.18 and 20.19 from Kister [23] illustrate the collected data superimposed on the SLE chart for the specified pressure drops, specific packing and column size, and packing heights at designated packing factors. Kister [23] recommends using the specific SLE chart (also see Kister [19] for a wide selection of charts) and interpolating and extrapolating the curves when the design/operating requirements fall close to the data points on the selected chart. Extrapolating too far can ruin the validity of the pressure drop results.



Basis: Fp-56
Pressure drop measured in inches H_2O/ft

Figure 20.18 SLE Data Chart for 1-in. metal Pall rings, aqueous systems, pressure drop only. Data from 15-84 in. dia. test columns with packed heights of 2-10 ft. Reproduced with permission of the American Institute of Chemical Engineers, Kister, H. Z. and Gill, D. R., Chemical Engineering Progress, V. 87, No. 2 © (1991) p. 32. All rights reserved.



Basis: Fp–27
Pressure drop measured in inches H_2O/ft

Figure 20.19 SLE Data Chart for 2-in. metal Pall rings, aqueous systems, pressure drop only. Data from 30-84 in. dia. test columns with packed heights of 2-18 ft. Reproduced with permission of the American Institute of Chemical Engineers, Kister, H. Z. and Gill, D. R., Chemical Engineering Progress, V. 87, No. 2 © (1991) p. 32. All rights reserved.

When the design/operating requirements are far from the chart's data points, contact the manufacturer of the packing for data and also consider selecting another type and/or size of packing to provide a better fit. In selecting a chart to use, do not overlook the physical nature of the process system, i.e., whether predominantly aqueous or non-aqueous, and whether it involves foaming characteristics or high or low viscosity fluids. Kister's assessment is that both the study in Reference 23 and Strigle [2] show that the large volume of published data does fit the Strigle [2] charts (Figure 20.13E, F) quite well and gives "good pressure drop predictions." A significant variation is that the curves predict an optimistic (too low) value for non-aqueous systems at high flow parameters of the chart. Also, similar optimistic values are noted for non-aqueous systems at low flow parameters, e.g., vacuum distillation. Unfortunately, most data were obtained on small scale as compared with industrial equipment, and so the statistical fit is still not adequate to use with confidence in industrial design [23].

20.20 Performance Comparisons

Strigle [24] presents some helpful comparisons with reference to Figure 20.14, and Tables 20.27 to 20.31 from [24] present valuable key design and operational points related to Figure 20.14.

For design, select C_s above the loading point C (on Figure 20.14) and below the maximum efficiency rate, point D. Usually, the flow at point C is 70 to 75% of that at maximum point F [24]. The design C_s (point E) should allow at least 15% increase in rates before reaching the maximum operating capacity of the packing. This allows for usual variation in operations [24]. The design C_s value should be at least 80 to 85% of the maximum operational C_s (MOC) for the best turndown condition during operation.

20.20.1 Prediction of Maximum Operating Capacity (MOC)

Strigle [24] proposed this term to better describe the performance of a packed column at or near the previously described "loading point." Kister [23] evaluated the limited published data and proposed using the MOC at 95% of the flood point. The flood point can be estimated by Equation 20.18 or from the plots in References 19 and 23. The data are reported to be within 15–20% of the prediction [23]. See Figure 20.14 for the identification of MOC on the HETP vs. C_s chart. For more accurate information, contact the respective packing manufacturers.

20.21 Capacity Basis for Design

Whether for a distillation, absorption, or stripping system, the material balance should be established around the top, bottom, and feed sections of the column. Then, using these liquid and vapor rates at actual flowing conditions,

Table 20.27 Maximum capacities of various packings.

Packings	C_s at maximum efficiency, ft/sec	ΔP at maximum efficiency H_2O/ft	C_s at maximum capacity, ft/sec	ΔP at maximum capacity, H_2O/ft
2 in. Pall Rings	0.295	0.81	0.315	1.29
2 in. Intalox Saddles	0.248	0.96	0.279	1.74
#50 Intalox Metal Packing	0.327	0.52	0.345	0.88
#40 Intalox Metal Packing	0.290	0.60	0.310	0.99
1-1/2-in. Pall Rings	0.269	0.95	0.287	1.48
1-1/2-in. Intalox Saddles	0.211	1.09	0.237	1.88
#40 Intalox Metal Packing	0.290	0.60	0.310	0.99
#25 Intalox Metal Packing	0.260	0.96	0.278	1.62

Note: #25 Intalox®, Norton = app. 1-in size.
#40 Intalox®, Norton = app. 11/2-in size.
#50 Intalox®, Norton = app. 2-in size.
Reproduced by permission of The American Institute of Chemical Engineers, Strigle, R. F., Jr. and Rukovena, F. *Chem. Eng. Prog.* Vol. 75, Mar. © (1979) p. 86, all rights reserved.

Table 20.28 Design efficiency and capacity for selected packings.

Packing	Design C_s, ft/sec	Design HETP, ft
2 in. Pall Rings	0.256	2.32
2 in. Intalox Saddles	0.216	2.50
#50 Intalox Metal Packing	0.284	2.12
#40 Intalox Metal Packing	0.252	1.74
1-1/2-in. Pall Rings	0.234	1.78
1-1/2-in. Intalox Saddles	0.183	1.87
#40 Intalox Metal Packing	0.252	1.74
#25 Intalox Metal Packing	0.226	1.38

Note: #25 Intalox®, Norton = app. 1-in size.
#40 Intalox®, Norton = app. 1 1/2-in size.
#50 Intalox®, Norton = app. 2-in size.
See Figure 9.22 for C_s vs. Tower internals.
Used by permission of The American Institute of Chemical Engineers, Stringle, R. F., Jr. and Rukovena, F. *Chem. Eng. Prog.*, Vol. 75 Mar. © (1979) p. 86, all rights reserved.

determine the flooding and maximum operating points or conditions. Then, using Figure 20.13B, 20.13E, or 20.13F, establish the pressure drop or assume a pressure drop and back-calculate a vapor flow rate, and from this a column diameter. Even though the column diameter may show a difference between the requirements at its top, bottom, and/or middle, it is not necessary to design a column with varying diameter to deal with these theoretically varying requirements. Sometimes the calculations show that a larger diameter in the upper section for vacuum columns would give better performance. A constant diameter column should be thoroughly evaluated before pursuing this, since it is the least expensive option and is somewhat easier to fit all the parts together. Even though the pressure drop or HETP may not be the same throughout [34–36], adjustments can be made, if warranted, even by changing the

Table 20.29 Comparison of maximum capacity designs.

Packing	Relative tower diameter	Relative packed height	Relative packing volume
2 in. Pall Rings	1.00	1.00	1.00
2 in. Intalox Saddles	1.09	1.08	1.28
#50 Intalox Metal Packing	0.95	0.91	0.82
#40 Intalox Metal Packing	1.01	0.75	0.76
1-1/2-in. Pall Rings	1.00	1.00	1.00
1-1/2-in. Intalox Saddles	1.13	1.05	1.34
#40 Intalox Metal Packing	0.96	0.98	0.91
#25 Intalox Metal Packing	1.02	0.78	0.80

Reproduced by permission: The American Institute of Chemical Engineers, Strigle, R. F., Jr., and Rukovena, F., *Chem. Eng. Prog.* Vol. 75, Mar., © (1979) p. 86, all rights reserved.

Table 20.30 Comparison of constant pressure drop designs.

$\Delta P = 0.5$ in H_2O per theoretical plate			
Packing	**Relative diameter**	**Relative height**	**Relative volume**
2 in. Pall Rings	1.00	1.00	1.00
2 in. Intalox Saddles	1.10	1.08	1.31
#50 Intalox Metal Packing	0.85	0.91	0.66
#40 Intalox Metal Packing	0.89	0.75	0.59
1-1/2- in. Pall Rings	1.00	1.00	1.00
1-1/2-in. Intalox Saddles	1.13	1.05	1.34
#40 Intalox Metal Packing	0.87	0.98	0.74
#25 Intalox Metal Packing	0.97	0.78	0.73

Reproduced by permission: The American Institute of Chemical Engineers, Strigle, R. F., Jr., and Rukovena, F., *Chem. Eng. Prog.* Vol. 75, Mar., © (1979) p. 86, all rights reserved.

Table 20.31 Generalized pressure drop rules.

Service	Design pressure drop in. in. H_2O/ft. of packed depth
Absorbers and Regenerators (Non-foaming System)	0.25–0.40
Absorbers and Regenerators	0.10–0.25
Atmospheric or Pressure Stills and Fractionators	0.40–0.80
Vacuum Stills and Fractionators	0.10–0.40

Source: Branan, C., The Process Engineer's Pocket Handbook, Gulf Publishing Co., Houston, Texas, (1976).

packing size or style/type in various segments of the column. This requires some careful calculations for the effective HETP and the total height of the individual sections and then the total column/tower. But it is often worth the effort, particularly for tall towers, say above 20–30 ft of packing. For low pressures as well as other columns, determine the pressure drop for each packed section of the column, plus the pressure drop through the internal components.

20.21.1 Flooding

In packed columns, flooding is caused when the energy potential of the vapor is greater than that of the liquid (i.e., when all the void spaces in the packing are filled with liquid), so that the liquid travels up and the column floods. This gives a ΔP of 2–3 in. H_2O/ft of packing depth, and a column cannot be operated successfully under these conditions. Flooding depends on a number of parameters, and empirical correlations of the variables have been presented by Bain and Hougen [37], Lobo *et al.* [29], Sherwood *et al.* [38], and Eckert [39].

Eckert's modification of Sherwood's flooding correlation has been widely applied for designing packed columns. Blackwell [40] developed mathematical models representing the generalized pressure drop curves excluding the flooding condition. Kessler and Wankat [41] formulated a model to represent Eckert's flooding curve within the range of the original data.

20.21.2 Operating and Design Conditions

Packed towers are typically operated at a gas velocity that corresponds to about 50–80% of flooding condition. This usually results in a ΔP of 0.5–1.0 in. H_2O/ ft of packing. Alternatively, a large decrease in flow rate causes channeling through the packed bed and could result in poor column performance.

Two principal design methods have often been employed in sizing packed towers. One is the selection of an allowable ΔP, and the other is to select some fraction of the flooding capacity. Nguyen [42, 43] uses the latter and employs Chen's least-square method to determine the tower diameter at 50% of the superficial flooding rate. This means that the ΔPs are 2 in. H_2O/ ft. of bed or greater. Here, an allowable ΔP is selected to determine the diameter of a packed tower. Kessler and Wankat's model is employed to determine the vapor rate at 70% of the flooding condition.

The maximum recommended design ΔP is 1.0 in H_2O/ft of liquids with a specific gravity of 0.8 or more. A lower maximum design ΔP is recommended for lighter liquids. The recommended design ΔP criterion for sizing a packed tower should fall within the ranges as shown in Table 20.32 [44]. Packed beds may be scaled up using the modified Sherwood correlation as illustrated in Figure 20.13C. The packing factor for various packing types is as shown in Tables 20.26A–F, because the capacity of the bed is independent of tower diameter and height. Packings that have sizes too large for the tower diameter will show a greater contact caused by the extra voidage, (i.e., where the pieces of packing contact the tower wall) and a lower ΔP than that predicted by their published packing factors. This effect reduces to zero for most packings when the ratio of packing size to diameter exceeds 10:1 [45]. This factor is necessary when data from a pilot plant tower, which is filled with large packings, are used for scale-up.

Design Equations

The packed tower sizing method is based upon the generalized ΔP correlation as shown in Figure 20.13C.

The abscissa on the generalized ΔP curve is expressed as

$$X = \left(\frac{L}{G}\right)\left[\frac{\rho_G}{(\rho_L - \rho_G)}\right]^{0.5} \tag{20.39}$$

where $0.01 < X < 2.0$ and the ordinate is expressed as

$$Y = \frac{\left(G_I^2 F_p \mu_L^{0.1}\right)}{[\rho_G(\rho_L - \rho_G)g]} \tag{20.40}$$

Table 20.32 Constants for each pressure drop correlation.

Pressure drop in. H_2O/ft. of packed depth	Constants	Pressure drop in. H_2O/ft. of packed depth	Constants
0.05	$C_0 = -6.30253$	0.50	$C_0 = -4.39918$
	$C_1 = -0.60809$		$C_1 = -0.99404$
	$C_2 = -0.11932$		$C_2 = -0.16983$
	$C_3 = -0.00685$		$C_3 = -0.00873$
	$C_4 = -0.00032$		$C_4 = -0.00343$
0.10	$C_0 = -5.50093$	1.0	$C_0 = -4.09505$
	$C_1 = -0.78508$		$C_1 = -1.00120$
	$C_2 = -0.13496$		$C_2 = -0.15871$
	$C_3 = -0.00134$		$C_3 = -0.00797$
	$C_4 = -0.00174$		$C_4 = -0.00318$
0.25	$C_0 = -5.00319$	1.5	$C_0 = -4.02555$
	$C_1 = -0.95299$		$C_1 = -0.98945$
	$C_2 = -0.13930$		$C_2 = -0.08291$
	$C_3 = -0.01264$		$C_3 = -0.03237$
	$C_4 = -0.00334$		$C_4 = -0.00532$

Source: Blackwell, W.W., Chemical Process Design On A Programmable Calculator, McGraw-Hill Book Company Inc., 1984.

where $0.001 < Y < 0.15$

$$Y = \exp[C_0 + C_1 \ln X + C_2(\ln X)^2 + C_3(\ln X)^3 + C_4(\ln X)^4] \tag{20.41}$$

and C_0–C_4 are constants derived for each ΔP. The values of these constants for each ΔP correlation are shown in Table 20.33. Table 20.33A shows simulation results at various pressure drops from 0.05-1.5 inch H_2O/ft. The results show a decrease in tower diameter with respective increase in the pressure drop.

$$G_1 = \left\{\frac{[Y\rho_G(\rho_L - \rho_G)]}{F_p\,\mu_L^{0.1}}\right\}^{0.5}, \ \frac{\text{lb}}{\text{ft}^2.\text{s}} \tag{20.42}$$

The tower area is

$$\begin{aligned} A &= \frac{G}{(3600)(G_1)}, \text{ft}^2 \\ &= \frac{\pi D^2}{4}, \text{ft}^2 \end{aligned} \tag{20.43}$$

The tower diameter is defined by

$$D = \left(\frac{4A}{\pi}\right)^{0.5}, \text{ft.} \tag{20.44}$$

Table 20.33 Input data and computer results of Example 20.1.

```
DATA201.DAT
3230.6          1136.0          0.0738
62.3            155.0           1.0             1.0
```

PACKED TOWER DESIGN	
Liquid Flow Rate, lb/hr	3230.6
Gas Flow Rate, lb/hr	1136.0
Gas Density, lb/ft^3	0.0737
Liquid Density, lb/ft^3	62.3
Tower Packing Factor	155.0
Pressure Drop, inch H_2O/ft:	1.0
Liquid Viscosity, cP	1.00
X Co-Ordinate in the Generalized Plot	0.0979
Y Co-Ordinate in the Generalized Plot	0.0719
Calculated Vapor Rate, lb/s ft^2	0.2617
X Co-Ordinate At Flooding Condition	0.0978
Y Co-Ordinate AT Flooding Condition	0.1335
Vapor Rate at 70% Flooding Condition, lb/s ft^2	0.2496
Vapor Velocity at 70% Flooding Condition, ft/s	3.39
Tower Area, ft^2	1.21
Tower Diameter, inches	14.87
Tower Diameter, ft.	1.24
Tower Area at 70% Flooding Condition, ft^2	1.67
Tower Diameter at 70% Flooding Condition, inches	17.49
Tower Diameter at 70% Flooding Condition, ft	1.46
Flood Pressure Drop, in H_2O/ft packing	3.926

At flooding condition, Eckert's flooding curve is represented as

$$\log\left(\frac{G_2^2 F_p \Psi \mu_L^{0.2}}{\rho_G \rho_L g}\right) \quad \text{vs.} \quad \log(F_{LG}) \tag{20.45}$$

where

$$F_{LG} = \left(\frac{L}{G}\right)\left(\frac{\rho_G}{\rho_L}\right)^{0.5} \tag{20.46}$$

Table 20.33A Computer results for ΔP of 0.05, 0.10, 0.25, 0.50, 1.0, and 1.5 in H_2O/ft.

Pressure drop ΔP, inch/H_2O	**0.05**	**0.10**	**0.25**	**0.5**	**1.0**	**1.5**
Liquid Flow Rate, lb/hr	3230.6	3230.6	3230.6	3230.6	3230.6	3230.6
Gas Flow Rate, lb/hr	1136.0	1136.0	1136.0	1136.0	1136.0	1136.0
Gas Density, lb/ft^3	0.0737	0.0737	0.0737	0.0737	0.0737	0.0737
Liquid Density, lb/ft^3	62.3	62.3	62.3	62.3	62.3	62.3
Tower Packing Factor	155.0	155.0	155.0	155.0	155.0	155.0
Liquid Viscosity, cP	1.00	1.00	1.00	1.00	1.00	1.00
X Co-Ordinate in the Generalized Plot	0.0979	0.0979	0.0979	0.0979	0.0979	0.0979
Y Co-Ordinate in the Generalized Plot	0.0043	0.0126	0.0273	0.0490	0.0719	0.1994
Calculated Vapor Rate, lb/s ft^2	0.0643	0.1097	0.1612	0.2161	0.2617	0.4359
X Co-Ordinate At Flooding Condition	0.0978	0.0978	0.0978	0.0978	0.0978	0.0978
Y Co-Ordinate AT Flooding Condition	0.1335	0.1335	0.1335	0.1335	0.1335	0.1335
Vapor Rate at 70% Flooding Condition, lb/s ft^2	0.2496	0.2496	0.2496	0.2496	0.2496	0.2496
Vapor Velocity at 70% Flooding Condition, ft/s	3.39	3.39	3.39	3.39	3.39	3.39
Tower Area, ft^2	4.90	2.88	1.96	1.46	1.21	0.72
Tower Diameter, inches	29.98	22.96	18.95	16.36	14.87	11.52
Tower Diameter, ft.	2.50	1.91	1.58	1.36	1.24	0.96
Tower Area at 70% Flooding Condition, ft^2	1.67	1.67	1.67	1.67	1.67	1.67
Tower Diameter at 70% Flooding Condition, inches	17.49	17.49	17.49	17.49	17.49	17.49
Tower Diameter at 70% Flooding Condition, ft	1.46	1.46	1.46	1.46	1.46	1.46
Flood Pressure Drop, in H_2O/ ft packing	3.926	3.926	3.926	3.926	3.926	3.926

Using Kessler and Wankat's model to represent the flooding condition,

$$X_1 = F_{LG} \tag{20.47}$$

and

$$Y_1 = \frac{\left(G_2^2 F_p \Psi \mu_L^{0.2}\right)}{(\rho_G \cdot \rho_L \cdot g_c)} \tag{20.48}$$

where

g_c = 32.174 lb_m./lb_f . ft/s^2
μ_L = liquid viscosity, cP
Ψ = ρ_{water}/ρ_L

At flooding condition,

$$\log(Y_1) = -1.6678 - 1.085 \log(X_1) - 0.29655(\log X_1)^2 \tag{20.49}$$

The mass vapor rate (lb/ft^2.s) at 70% flooding condition can be expressed as

$$G_2 = 0.7\left(\frac{Y_1 \cdot \rho_G \cdot \rho_L \cdot g_c}{F_p \cdot \Psi \cdot \mu_L^{0.2}}\right)^{0.5}, \left(\frac{lb}{s.ft^2}\right) \tag{20.50}$$

The vapor velocity (ft/s) at 70% flooding condition is expressed as

$$V_{GF} = \frac{G_2}{\rho_G}, ft/s \tag{20.51}$$

Since the flooding correlation is not perfect, a 95% confidence with a safety factor of 1.32 is used to determine the area at flooding condition (Bolles and Fair [46]).

Therefore, area of the tower at flooding condition is

$$Area_{flood} = 1.32 \times \left[\frac{G}{(3600 \times G_2)}\right], ft^2 \tag{20.52}$$

The tower diameter at 70% flooding condition is

$$dia_{flood} = \left[\frac{4 \times Area_{flood}}{\pi}\right]^{0.5}, ft. \tag{20.53}$$

In the region below the flooding curves, the pressure drop can be correlated with an equation of the form

$$\Delta P = \alpha\left(10^{\beta L_1}\right)\left(\frac{G_1^2}{P_G}\right) \tag{20.54}$$

The pressure drop at incipient flooding from Kister and Gill's [22, 23] equation is

$$\Delta P_{flood} = 0.115 F_p^{0.7}, (in.\, H_2O/ft.\, of\ packing) \tag{20.21}$$

where

A = Tower area, ft^2, Equation 20.43
$Area_{flood}$ = Tower area at 70% flooding condition, ft^2, Equation 20.52
C_0–C_4 = Pressure drop correlation constants.
D = Packed tower diameter, ft.
dia_{flood} = Tower diameter at 70% flooding condition, ft, Equation 20.53
F_{LG} = Flow parameter in Equation 20.46
F_p = Packing factor for packing material
G = Vapor rate, lb/h.
G_1 = Vapor mass velocity, lb/(ft^2.s. tower cross section), Equation 20.42
G_2 = Vapor mass velocity, lb/(ft^2.s. tower cross section), Equation 20.50
g_c = conversion factor, 32.174 lb_m./lb_f . ft/s^2
L = Liquid rate, lb/h

L_1 = Liquid mass velocity, lb/(ft^2 sec. tower cross-section)
X = Abscissa in the generalized pressure drop correlation chart
Y = Ordinate in the generalized pressure drop correlation chart
Y_1 = Ordinate in Eckert's flooding curve
V_{GF} = Vapor velocity at 70% of flooding condition, ft/sec., Equation 20.51.

Greek letters

α, β = Constants in Tables 20.23 and 20.24
μ_L = Liquid viscosity, cP
ρ_L = Liquid density, lb/ft^3
ρ_G = Vapor density, lb/ft^3
Ψ = Ratio of water density to liquid density, ρ_{water}/ρ_L
υ = Kinematic liquid viscosity, centistokes, note: Centistokes = centipoise/(ρ_L/62.4)
ΔP_{flood} = Pressure drop at incipient flooding, in. H_2O/ft. of packing.

Example 20.1

A packed tower with 1-inch Raschig rings is to be used to absorb ammonia from air by contacting the gas stream with water. The entering gas flow is 40 lb mol/h and contains 5.0% of NH_3. Ninety percent of the NH_3 is to be removed from the air. The entering water flow rate is 3,200 lb/h. Absorption is to be carried out at 1 atm. Abs. and 20°C. Estimate

1. The diameter of the tower for a ΔP of 1.0 in. H_2O/ft. packing.
2. The vapor rate at 70% of flooding condition.
3. Determine the diameters for pressure drops of 0.05, 0.10, 0.25, and 1.5 in.

Physical properties:

$$\rho_G = (28.4)\left(\frac{1}{259}\right)\left(\frac{273}{293}\right)$$
$$= 0.0737 \text{ lb/ft}^3$$

$$\rho_L = 62.3 \text{ lb/ft}^3$$

$$\mu_L = 1 \text{ cP}$$

$$F_p = 155$$

$$\Psi = 1.0$$

Solution

When calculating the tower diameter, conditions at its base should be used because this corresponds to the maximum flow rates.

At the bottom of the column:

$$MW_{gas} = (0.05)(17) + (0.95)(29)$$
$$= 28.4 \text{ lb/lb mol.}$$

Molecular weight of ammonia, $MW_{NH_3} = 17 \dfrac{\text{lb}}{\text{lb mol}}$

NH_3 removed from the air = (0.9) (0.05) (40)

$$= \left(1.8 \frac{\text{lb mol}}{\text{h}}\right)\left(17 \frac{\text{lb}}{\text{lb mol}}\right)$$

$$= 30.6\ \text{lb/h}.$$

$$\text{Gas flow rate at the bottom} = \left(40 \frac{\text{lb mol}}{\text{h}}\right)\left(28.4 \frac{\text{lb}}{\text{lb mol}}\right)$$

= 1,136 lb/h

Liquid flow rate at the bottom = 3200 + 30.6
= 3230.6 lb/h

Using the generalized pressure drop correlation of Strigle (Figure 20.13F).

$$\text{The flow parameter, } F_{LG} = \frac{L}{G}\sqrt{\frac{\rho_G}{\rho_L}}$$

$$= \frac{3230.6}{1136.0}\left[\frac{0.0737}{62.3}\right]^{0.5} = 0.0978$$

From Figure 20.13F at $\Delta P = 1.0\,\text{in.}H_2O/\text{ft}$, the ordinate is 0.069.

That is,

$$0.069 = \frac{G_1^2 F \nu^{0.1}}{32.2\rho_G[\rho_L - \rho_G]}$$

The liquid viscosity in centistokes, $\nu = \text{cP}/(\rho_L/62.4)$
= 1/(62.3/62.4) = 1.0016 cSt.

and the mass vapor rate G_1 is expressed as

$$G_1 = \left[\frac{0.069 \cdot 32.2 \cdot \rho_G \cdot (\rho_L - \rho_G)}{F_p \nu^{0.1}}\right]^{0.5}, \text{lb/s.ft}^2$$

$$= \left[\frac{(0.069)(32.2)(0.0737)(62.3 - 0.0737)}{(155)(1.0016)^{0.1}}\right]^{0.5}$$

$$= 0.2564\ \text{lb/s.ft}^2$$

$$\text{Area of the packed tower, Area} = \frac{G}{G_1} = \frac{1136}{(0.2564)(3600)}\left\{\frac{\cancel{\text{lb}}}{\cancel{\text{hr}}} \cdot \frac{1}{\dfrac{\cancel{\text{lb}}}{\text{ft}^2 . \cancel{\text{sec}}}} \cdot \frac{\cancel{\text{hr}}}{\cancel{\text{sec}}}\right\}$$

Area = 1.23 ft^2

$$\text{Tower diameter, } d = \left(\frac{4 \cdot \text{Area}}{\pi}\right)^{0.5}, \text{ ft.}$$

$$= \left(\frac{4 \times 1.2308}{\pi}\right)^{0.5}$$
$$= 1.25 \text{ ft. (15 inches).}$$

(b). Using Kessler and Wankat's correlation to represent the flooding condition [41],

$$X_1 = F_{LG}$$
$$= 0.0978$$

and

$$Y_1 = \frac{\left(G_2 F_p \Psi \mu_L^{0.2}\right)}{(\rho_G \cdot \rho_L \cdot g)}$$

At flooding condition:

$$\log(Y_1) = -1.6678 - 1.085 \log(X_1) - 0.29655(\log X_1)^2$$
$$= -1.6678 - 1.085 \log(0.0978) - 0.29655 (\log 0.0978)^2$$
$$= -0.8746$$
$$Y_1 = 0.13346$$

The mass vapor rate (lb/ft^2.s) at 70% flooding condition can be expressed as

$$G_2 = 0.7\left(\frac{Y_1 \cdot \rho_G \cdot \rho_L \cdot g}{F_p \cdot \Psi \cdot \mu_L^{0.2}}\right)^{0.5}, \left(\frac{\text{lb}}{\text{s} \cdot \text{ft}^2}\right)$$
$$= 0.7\left[\frac{(0.1335)(0.0737)(62.3)(32.2)}{(155)(1)(1)^{0.2}}\right]^{0.5}$$
$$= 0.2497 \text{ lb/s. ft}^2$$

The vapor velocity (ft/s) at 70% flooding condition is

$$V_{GF} = \frac{G_2}{\rho_G}\left\{\frac{\cancel{\text{lb}}}{\text{s.}\cancel{\text{ft}^2}} \cdot \frac{1}{\frac{\cancel{\text{lb}}}{\cancel{\text{ft}^3}}}\right\}$$
$$= 0.2497/0.0737$$
$$= 3.389 \text{ ft./s}$$

The tower area at 70% flooding condition using a safety factor of 1.32 is

$$
\begin{aligned}
\text{Area}_{\text{flood}} &= 1.32\frac{G}{(3600)G_2}, \quad \left\{\frac{\cancel{\text{lb}}}{\cancel{\text{hr}}}\cdot\frac{1}{\dfrac{\cancel{\text{lb}}}{\cancel{\text{s}}.\text{ft}^2}}\cdot\frac{\cancel{\text{hr}}}{\cancel{\text{s}}}\right\} \\
&= 1.32\frac{1136}{(3600)(0.2497)} \\
&= 1.66\ \text{ft}^2
\end{aligned}
$$

Tower diameter is

$$
\begin{aligned}
\text{dia},_{\text{flood}} &= \left[\frac{4\times 1.66}{\pi}\right]^{0.5} \\
&= 1.46\ \text{ft.}(17.5\ \text{in.})
\end{aligned}
$$

The pressure drop at incipient flooding from Kister and Gill's [22, 23] equation is

$$
\begin{aligned}
\Delta P_{\text{flood}} &= 0.115 F_p^{0.7}, (\text{in. } H_2O/\text{ft. of packing}) \\
&= 0.115(155)^{0.7} \qquad (20.18) \\
&= 3.926\ \text{in. } H_2O/\text{ft. of packing.}
\end{aligned}
$$

Computer Program

A computer program PROG 201.for has been developed that calculates the diameter of a packed tower for pressure drops of 0.05, 0.10, 0.25, 0.50, 1.0, and 1.5 in. H_2O/ft. The simulation is based on the generalized pressure drop correlation of Sherwood *et al.* [38]. The curves have been fitted to a series of equations developed by Blackwell [40]. In addition, the program calculates the vapor rate and vapor velocity at 70% of flooding condition based on Kessler and Wankat's correlation and uses the safety of 1.32 to determine the diameter of the column at 70% flooding condition. The pressure drop at flooding condition is determined using Kister and Gill's equation [22, 23]. Table 20.34 shows the simulation results of the packed tower at varying pressure drops for Example 20.1.

Table 20.34 Nutter ring hydraulic coefficients.

Nutter ring (TM) size	**C_1**	**X_1**	**X_2**	**X_3**	**X_4**
0.7	0.141	0.735	−0.00646	1.0	0.05
1	0.095	0.870	−0.00562	1.10	0.08
1.5	0.070	0.969	−0.00230	1.12	0.08
2	0.059	1.000	0.0	1.15	0.08
2.5	0.051	1.0394	0.000653	1.15	0.08
3	0.037	1.124	0.002076	1.15	0.10

Used by permission of Nutter Engineering, a Harsco Corp.

20.22 Proprietary Random Packing Design Guides

Norton Intalox® Metal Tower Packing (IMTP®)

Norton offers a new high-performance system centered around an improved Intalox® metal tower packing including effective internals to provide the distributor and pressure drop consistent with the higher performance of the packing itself.

Figures 20.6J and 20.6U illustrate the IMTP packing. The manufacturer's key performance descriptions are [47]

1. Greater capacity and efficiency than fractionation trays and other dumped packings.
2. Pressure drop approximately 40% lower than equivalent size Pall rings.
3. Low liquid hold-up.
4. Structural strength allows packing depths to 50 ft or more.
5. Easy to use in distillation, from deep vacuum where low pressure drop is beneficial, to high pressure where capacity surpasses many trays.
6. HETP values nearly independent of flow rate.
7. Packing properties related to a performance curve:
   - **(a)** The system base HETP of a packing, which is the flat HETP value produced by uniform distribution.
   - **(b)** The efficient capacity (not the same as hydraulic capacity or flood point) of a packing, which is the greatest vapor rate at which the packing still maintains the system base HETP. Norton [33] rates packings by percent of efficient capacity rather than percent of flood.

Capacity Correlation [47]

Figures 20.13G and 20.13I present the proprietary estimating capacity charts for the various sizes of the Intalox® packing for a non-foaming system. The system base HETP of a packing is the flat HETP value produced by uniform distribution; see Figure 20.14.

The terminology for the chart referenced to Norton's [47] Intalox® random packing of various sizes designated as

Size No.	Dimensions, in.
15	–
25	1
40	–
50	2
70	–

Design information used by permission of Norton Chemical Process Products Co.

$$\text{Flow parameter, } X = L/G\sqrt{\rho_G/\rho_L}\text{, chart} \tag{20.55}$$

$$\text{Capacity parameter, } C_s = V\sqrt{\rho_G/\rho_L - \rho_G}\text{, ft/sec. or m/sec} \tag{20.56}$$

Capacity rating, C_o = feet/sec., or m/sec, from chart

$$\text{Efficiency capacity, } C_{sc} = C_o\left[\frac{\sigma}{20}\right]^{0.16}\left[\frac{\mu}{0.2}\right]^{-0.11}\text{, ft/sec/ or meters/sec} \tag{20.57}$$

This is the greatest vapor rate at which the packing still maintains the system base HETP [47].

Capacity rating = $[C_s/C_{sc}]$ (100) = % capacity [47].

Pressure drop equation: For IMTP packing, non-foaming system, use Figure 20.13G or 20.13I.

where flow parameter, X = as previously noted, Equation 20.55.

$$\text{Capacity parameter, } Y = FC_s^2\nu^{0.1} \tag{20.58}$$

Value for F coefficient (used by permission, Norton [47])

IMTP size	**No. 15**	**No. 25**	**No. 40**	**No. 50**	**No. 70**
F when C_s in m/sec	549	441	258	194	129
F when C_s in ft/sec	51	41	24	18	12

where
ν = liquid kinematic viscosity, centistokes
= μ/(sp.gr.) centistokes
μ = liquid viscosity, centistokes
σ = surface tension, dynes/cm
L = liquid mass rate, lb/h
G = gas mass rate, lb/h
ρ_L = liquid density, lb/ft^3
ρ_G = gas density, lb/ft^3
V_g = V = superficial gas velocity, ft/s or m/s

Note: 1 in. water/ft = 1.87 mm Hg/ft = 6.15 mm Hg/m

$$\text{Superficial vapor velocity, } V = G/(\rho_G A) \tag{20.59}$$

where $A = (\pi/4)D^2$, ft^2

20.22.1 Packing Efficiency/Performance for IMTP Packing [33]

This is usually expressed as HETP, and for the IMTP packing when a high-performance (uniform) liquid distributor is used in a column, the HETP is independent of the tower diameter and packing depth. Norton [47] has developed a concept for evaluating HETP for fluids of the types (a) non-aqueous, (b) non-reacting and non-ionizing, and (c) low relative volatility (less than three) [47], as follows:

System base HETP of IMTP packing for distillation and reboiled stripping:

$$\text{(A) System Base HETP} = A\left[\frac{\sigma}{20}\right]^{-0.16}(1.78)^{\mu}, \text{For } \mu \le 0.4\text{cP} \tag{20.60}$$

$$\text{(B) System Base HETP} = B\left[\frac{\sigma}{20}\right]^{-0.19}\left[\frac{\mu}{0.2}\right]^{0.21}, \text{For } \mu > 0.4\,\text{cP} \tag{20.61}$$

expressed as millimeters or inches, depending on constants used for A and B in the previous equations. If $\sigma > 27$, use $\sigma = 27$, dynes/cm; μ = liquid viscosity, cP. Use the values of A and B as follows [Reference 47 by permission]:

Values of A and B for Equations 20.60 and 20.61					
IMTP Size	**No. 15**	**No. 25**	**No. 40**	**No. 50**	**No. 70**
A-mm	271.3	331.4	400.7	524.9	759.9
A-in	10.68	13.05	15.78	20.67	29.92
B-mm	296.7	366.1	438.3	579.9	831.2
B-in	11.68	14.41	17.25	22.83	32.72

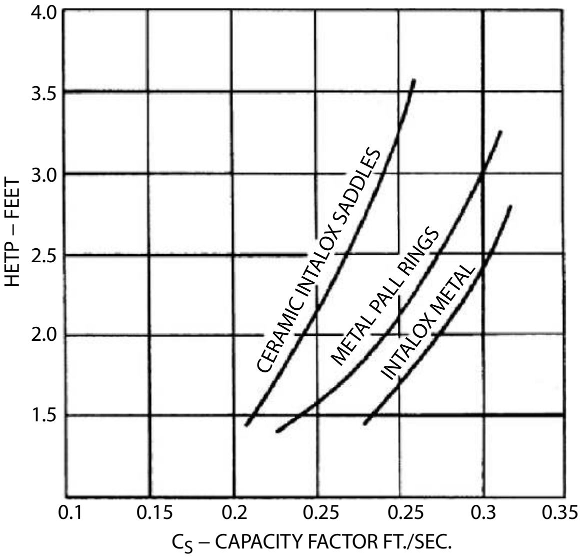


Figure 20.20 Packing comparison at design point for Intalox® ceramic saddles, metal pall rings, and metal Intalox®. Reproduced by permission of American Institute of Chemical Engineers, Strigle, R. F., Jr., and Rukovena, F., Chemical Engineering Progress, Mar. © (1979) p. 86. All rights reserved.

For commercial towers with good liquid/vapor distribution, Norton [33] recommends standard designs use HETP values 13% above the system base HETP. If the system under consideration does not meet the physical properties limit, either use a conservative estimate or use actual plant or published data for the system. For comparison of HETP values for selected packings, see Strigle and Rukovena [24], Figure 20.20.

Example 20.2: Hydrocarbon Stripper Design (Figure 20.21)

Design a packed tower splitter for a light hydrocarbon plant. The conditions of operation as determined from material balance are as follows:

Operate at 430 psig, feed from ethylene tower at purification unit.

Feed:

Volume:	55% ethylene
	45% ethane

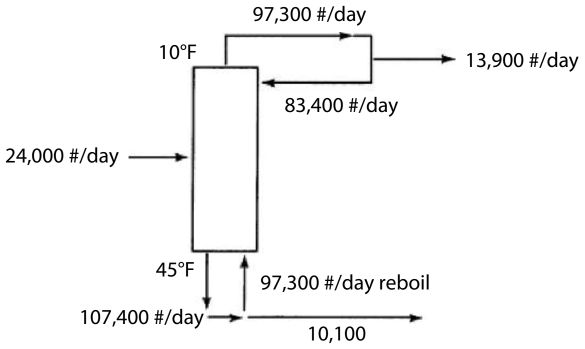


Figure 20.21 LHC plant stripper using packed tower for Example 20.2.

Specifications to be:

Overhead:	90% ethylene
Bottoms:	95% ethane
Feed:	13,000 lb/day ethylene
	11,000 lb/day ethane

Determine the feed temperature from ethylene tower.

Solution

Bottoms: 25°F, Boiling point feed

Bubble point of bottoms: At 445 psia

Composition x_B		K @ 47°F	Kx	K @ 45F	Kx
Ethylene	0.05	1.36	0.068	1.35	0.067
Ethane	0.95	1.0	0.95	0.97	0.921
	1.00			1.018	0.988

Therefore, bubble point = 46°F

$$\alpha_{\text{bottoms}} = 1.35/0.98 = 1.38$$

Dew point on overhead: at 445 psia

Composition	y_{OH}	K@15°F	y/K
Ethylene	0.90	1.05	0.857
Ethane	0.10	0.75	0.138
	1.00		0.995

Dew point on overhead; 15°F

Bubble point on overhead: Try 10°F at 445 psia

Composition	x_D	**K**	**Kx**
Ethylene	0.90	1.00	0.90
Ethane	0.10	0.80	0.08
	1.00		0.98

Bubble point > 10°F < 15°F ≅ 12°F

$$\alpha_{Top} = 1.05/0.73 = 1.44$$

$$\alpha_{avg} = \sqrt{\alpha_{top}\alpha_{bottom}} = \sqrt{(1.38)(1.44)} = 1.41$$

Minimum Trays at Total Reflux

Fenske Equation:

$$\begin{aligned} N+1 &= \frac{\log(x_{lk}/x_{hk})_D(x_{hk}/x_{lk})_B}{\log\alpha_{avg}} \\ &= \frac{\log(0.90/0.10)(0.95/0.05)}{\log\ 1.41} \\ &= 15 \text{ theoretical trays} \end{aligned}$$

Minimum Reflux

Gilliland Plot:

$$\begin{aligned} (L/D)_{Min} &= \frac{x_D[1+(\alpha-1)x_F]-\alpha x_F}{(\alpha-1)(x_F)(1-x_F)} \\ &= \frac{0.90[1+0.41(0.55)]-(0.55)(1.41)}{(0.41)(0.55)(0.45)} = 3.23 \end{aligned}$$

Use actual L/D = 6.1

Theoretical Plates vs. Reflux

Gilliland Plot

(L/D)	**Theoretical Plates**
3.25	Infinity
4.0	28.4
5.0	22.4
6.0	20.0
7.0	18.6
8.0	17.8
Infinity	14.2

Figure 20.21 shows the schematic diagram of loading, and Figure 20.22 shows the plot for this example, respectively.

From the material balance using (L/D) = 6.0; theoretical trays = 20

(a) Feed: 24,000 lb/day
(b) Distillate product: 13,900 lb/day
(c) Reflux into column: 83,400 lb/day @ 6:1 reflux, L/D
(d) Total gross overhead: 97,300 lb/day
(e) Total bottoms out: 10,100 lb/day

Top of column

$$\rho_v = \frac{(28)(445)(492)}{(359)(14.7)(470)} = 2.47\ \text{lb/ft}^3 @ 430\ \text{psig} \,\&\, 10°\text{F}$$

$$\rho_L = 0.39\ (62.4) = 24.3\ \text{lb/ft}^3$$

Liquid viscosity = 0.07 cP

From Figure 20.13C,

$$\frac{L}{G}\sqrt{\frac{\rho_V}{\rho_L}} = \frac{83{,}400}{97{,}300}\sqrt{\frac{2.47}{24.3}} = 0.27$$

Read:

$$\frac{G^2 F \mu^{0.1}}{\rho_G(\rho_L - \rho_G)g_c} = 0.067\ (\text{flooding avg.})$$

For average loading condition, read ordinate = 0.030.

For average flooding: (from chart) using 1-in. metal Pall rings with F = 56.

$$0.067 = G^2\ (56)\ (0.07)^{0.1}/((32.2)\ (24.3 - 2.47)\ (2.47))$$

$$G = 1.648\ \text{lb/s.ft}^2\ \text{for average flooding}$$

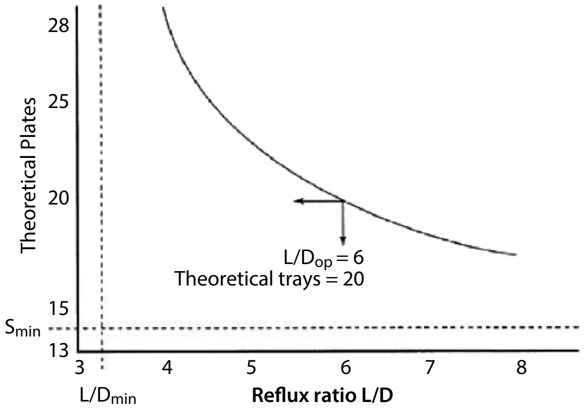


Figure 20.22 Gilliland plot for Example 20.2.

Loading velocity, $V_{\text{loading}} = 1.648/2.47 \left\{\dfrac{\text{lb}}{\text{sec.ft}^2} \cdot \dfrac{\text{ft}^3}{\text{lb}}\right\} = 0.667$ ft/s

Volume flow rate = 97,300/(24) (3,600) (2.47) = 0.4559 ft^3/s

Area required = $0.4559/0.667 \left\{\dfrac{\cancel{\text{ft}}^3}{\cancel{\text{sec}}.} \cdot \dfrac{\cancel{\text{sec}}}{\cancel{\text{ft}}}\right\} = 0.6835\ \text{ft}^2\ (= 98.4\ \text{in}^2)$

Column diameter, $D = \sqrt{\dfrac{4\,\text{Area}}{\pi}} = \sqrt{\dfrac{4 \times 0.6835}{\pi}} = 11.19$ in.

Use = 16-in. I.D. column

Bottom of column:

$$\mu_v = 0.0095 \text{ cP}$$

$$\mu_L = 0.07 \text{ cP}$$

$$\rho_v = \frac{30(445)(492)}{359(14.7)(505)} = 2.47 \text{ lb}/\text{ft}^3$$

$$\rho_L = 0.4\ (62.4) = 25 \text{ lb/ft}^3$$

$$\frac{L}{G}\sqrt{\frac{\rho_V}{\rho_L}} = \frac{107{,}400}{97{,}300}\sqrt{\frac{2.47}{25}} = 0.347$$

Using packing factor, F = 56 for 1-in. Pall metals rings

Reading chart:

$$\frac{G^2 F \mu^{0.1}}{\rho_G(\rho_L - \rho_G)g_c} = 0.057, \text{ flooding, for reference}$$

loading reads = 0.036

$$G^2(56)(0.07)^{0.1}/(2.27)(25 - 2.47)(32.2) = 0.057$$

$$G^2(0.02394) = 0.057$$

$$G = 1.5425 \text{ lb/s ft}^2.$$

Loading velocity, $V_{\text{loading}} = 1.5425/2.47 \left\{\dfrac{\cancel{\text{lb}}}{\text{sec.}\cancel{\text{ft}}^2} \cdot \dfrac{\cancel{\text{ft}}^3}{\cancel{\text{lb}}}\right\} = 0.6245$ ft/s

Volume flow rate = 97,300/(24) (3,600) (2.47) = 0.4559 ft^3/s

Area required = $0.4559/0.6245 \left\{\dfrac{\cancel{\text{ft}}^3}{\cancel{\text{sec}}.} \cdot \dfrac{\cancel{\text{sec}}}{\cancel{\text{ft}}}\right\} = 0.7299\ \text{ft}^2\ (= 105.1\ \text{in}^2)$

Column diameter, $D = \sqrt{\dfrac{4\,\text{Area}}{\pi}} = \sqrt{\dfrac{4 \times 0.7299}{\pi}} = 0.964$ ft. (11.56 in.)

Use 16-in. diameter

It is better to use smaller packing (less than 1 in.) in this diameter column (16-in. I.D.), such as ⅝ in. if recalculated above. Packing of ½ in. or ⅜ in. is better in this size column; however, the effects of packing factor should be calculated if changed from ⅝ in. packing.

HETP: Use HETP = 18 in. based on on-site column data and manufacturer's confirmation that for ⅝ in. Pall rings in this system, the 18 in. HETP should perform satisfactorily. Note: For each design, verify expected HETP through the manufacturer.

Redistribute the liquid every 10 ft and allow 2 ft additional for redistribution = 36 ft. Then use 4–10 ft sections of packing. For 20 theoretical plates, total performance packed height 20 (18 in./12 in.) = 30 ft. Allow for loss of equilibrium at (1) reflux entrance = 1 HETP, (2) feed entrance = 2 HETP, and (3) redistribution (2) = 2 HETP; totals 5 HETP.

$$5 \times 1.5 \text{ ft} = 7.5 \text{ ft extra packing}$$

Total packing to install = 30 ft + 7.5 ft = 37.5 ft; Use 40 ft Feed point, based on Kirkbride (see Chapter 19):

$$\log\frac{N_N}{N_M} = 0.206\log\left[\left(\frac{B}{D}\right)\left(\frac{x_{hF}}{x_{lF}}\right)\left(\frac{x_{lB}}{x_{hD}}\right)^2\right] \tag{20.62}$$

$$\begin{aligned}\log\frac{N_N}{N_M} &= 0.206\log\left(\frac{10{,}100}{13{,}900}\right)\left(\frac{0.45}{0.55}\right)\left(\frac{0.05}{0.10}\right)^2\\ &= 0.206\log 0.149\\ &= -0.206(0.837)\\ &= -0.1725\\ \log\frac{N_N}{N_M} &= 0.1725\end{aligned}$$

$$N_M/N_N = 1.487$$

$$N_M + N_N = 40 \text{ ft}$$

$$1.487\,(N_N) + N_N = 40$$

then, N_N = 16.08 ft, rectifying section; use 1 16.5-ft section
N_M = 23.92 ft, stripping section; use 2 12-ft sections

Allow 2 ft space between support of top section and redistributor of section 2 (from top).

Estimated pressure drop of loading from Figure 20.13C: = 0.40 in H_2O/ft of packing. For total packed height, $\Delta P \cong (0.40)(16.5 + 24) = 16.2$ in. H_2O total (packing only).

Preliminary evaluation of condenser requirements: Condense: 97,300 lb/day

$$L_v = 115 \text{ Btu/lb}$$

$$q = 97{,}300\,(115)/24 = 466{,}229 \text{ Btu/h}$$

For approximate estimate: assume overall heat transfer coefficient,

$$U = 100 \text{ Btu/hr (ft}^2\text{) (}^\circ\text{F)}$$

$$\Delta T = 35^\circ\text{F for (}-25^\circ\text{F propylene)}$$

$$\text{Area Estimated} = Q/U\,\Delta T$$

$$= 466{,}229/(100)(35)$$

$$= 133 \text{ ft}^2 \text{ (outside tubes)}$$

Recommend kettle type condenser.

Nutter Ring [48]

Nutter offers an improved high-performance random packing identified as Nutter Ring^{TM}; see Figure 20.6K. To achieve the best performance from any random packing, the liquid distributor must be level, and the distributor points of the discharging liquid to the packing must be uniformly distributed; see earlier discussion on this topic.

The manufacturer's key performance descriptions and claims are as follows:

1. Comparative tests at Fractionation Research, Inc. (FRI) [34] showed the No.2 Nutter ring^{TM} to have greater usable efficiency-capacity profile, with substantially lower pressure drop than is achieved with slotted rings. Usable capacity is comparable to 3½-in. Pall rings (see Figures 20.23A–C).
2. Ring form provides multiple circular crimped strips to encourage liquid surface renewal.
3. Perforated central trough enhances lateral liquid spreading and effectively wets the outside surface facing vapor flow.
4. Heavily ribbed main element for a high strength-to-weight ratio.
5. A pair of tapered slots and hoops provides maximum randomness with minimal nesting.
6. Efficiency enhanced by item (2) above, and the No.2 Nutter Ring is better than 2-in. Pall rings [36, 49].
7. Superior surface utilization in mass and heat transfer, allowing shorter packed bed heights. Turndown performance is superior over 2- and 3½-in. Pall rings.
8. Strength-to-weight ratio allows bed heights of 50 ft.
9. Lower pressure drop and greater usable capacity allow use of smaller diameter columns. Pressure drop per stage is 30–50% less than either size Pall ring.
10. Reproducible performance assured through uniform randomness [35].
11. All rings in sizes No. 7 to No. 3 diameter are proportional on all dimensions for accuracy in scale-up of pressure drop, capacity, and efficiency. See Figure 20.6K.
12. Plastic Nutter Rings^{TM} are rigid and energy efficient, and permit applications to produce pressure drops per theoretical stage and bed heights, not attainable with other random particle packing.

Capacity Correlation [50]

This design presentation is proprietary to Nutter Engineering and is definitely more accurate than applying the Nutter Ring^{TM} to the generalized pressure drop correlations (Figures 20.13B to 20.13F). Figures 20.13G and 20.13I do not apply because they are proprietary to another manufacturer. The procedures to follow supersede the equations in Nutter and Perry [50, 51] and are used by permission of Nutter Engineering, a Harsco Corp.

Determine % useful capacity; assume a column diameter, or calculate using an existing column under study. Usable capacity is defined as the maximum loading condition where efficiency does not deteriorate significantly from that achieved over a lower range of loadings [51]. For the Nutter ring, the limiting pressure drop for usable capacity is 1 in. of hot liquid for low-pressure systems. Designs should not exceed this value. This has been shown to hold for C_6–C_7 system at 5 psia and for hydrocarbons at atmospheric to 24 psia (see Figures 20.23A and 20.23B) [50].

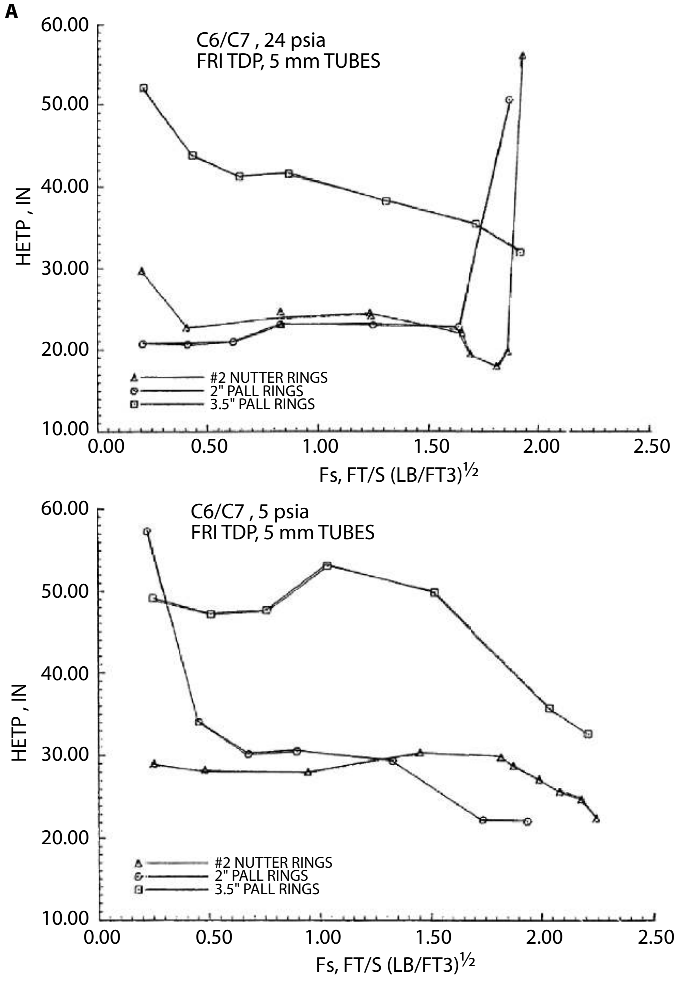


Figure 20.23A Comparison of HETP for No. 2 Nutter Rings™ and Pall rings in a C_6/C_7 system at 24 psia and 5 psia using the FRI tubed drip pan distributor. Data prepared and used by permission of Nutter Engineering, Harsco Corp. and by special permission of Fractionation Research, Inc. All rights reserved.

$$\% \text{ Useful capacity} = (100)\,(C_s)\,(F_c/C_2) \tag{20.63}$$

where

C_s = vapor rate = $V_s\,[\rho_v/(\rho_l - \rho_v)\,]^{0.5}$, ft/s (20.64)

ρ_v = vapor/gas phase density, lb/ft^3

ρ_l = liquid phase density, lb/ft^3

V_s = vapor superficial velocity, ft/s

C_2 = wet pressure drop intercept coefficient

= $(F_{Q1})\,(F_{size})\,(F_{system})$ (20.65)

F_{Q1} = liquid rate factor for C_2

F_{size} = size factor for C_2

F_{system} = physical properties factor for C_2

F_c = useful system capacity factor

= $(0.535)\,((\rho_l - \rho_v)/\,\rho_v)^{0.12}$ (20.66)

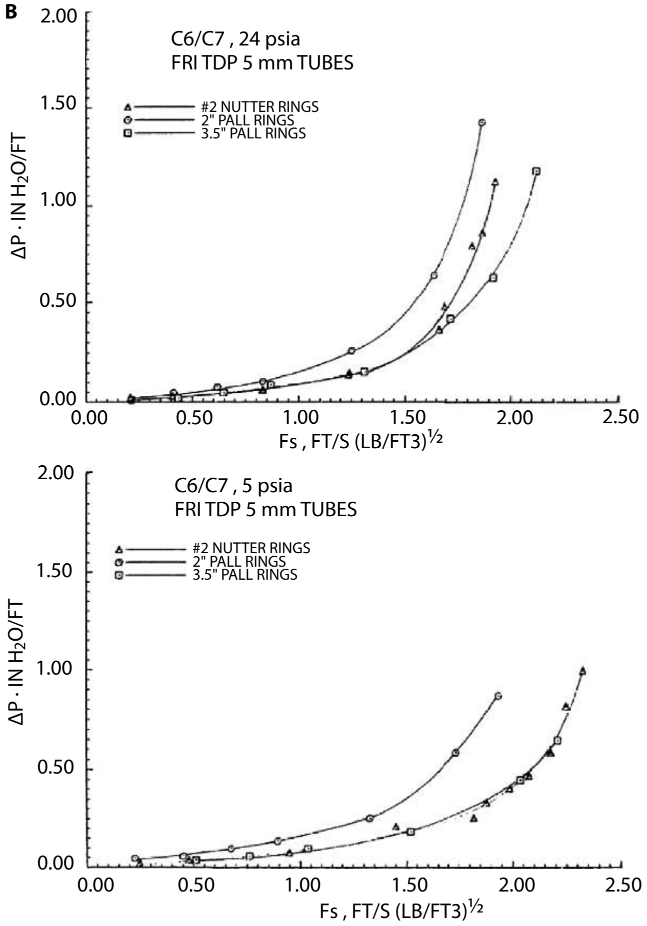


Figure 20.23B Comparison of HETP for No. 2 Nutter Rings™ and Pall rings in a C_6/C_7 system at 24 psia and 5 psia using the FRI tubed drip pan distributor. Data prepared and used by permission of Nutter Engineering, Harsco Corp. and by special permission of Fractionation Research, Inc. All rights reserved.

Limits: $0.728 <= F_c <= 1.04$

$$F_{Q1} = 0.428 - 0.0141Q_1 + 0.000326Q_1^2 - 3.7\times10^{-6}Q_1^3 + 1.47\times10^{-8}Q_1^4 \qquad (20.67)$$

Q_l = liquid superficial velocity, gpm/ft^2
F_s = vapor rate, $V_s\,(\rho_v)^{0.5}$, (ft/sec) (lb/ft^3)$^{0.5}$ (20.68)
F_{size} = $X_1 + X_2\,(Q_l)$ (20.69)
X_1, X_2 = constants from Table 20.34
F_{system} = $1.130\,(\sigma/\rho_l)^{0.179}$ (20.70)
σ = surface tension, liquid, dynes/ cm

$$\%\text{ System limit vapor velocity} = 100\,V_s/\left[(0.760)(\sigma/\rho_v)\right]^{0.461} \qquad (20.71)$$

not applicable when $((\rho_l - \rho_v)/\rho_v)^{0.5} > 4.5$

Pressure drop: dry bed

$$\Delta P_d = C_1\,(F_s)^2 = \text{dry bed pressure drop. in. water/ft} \qquad (20.72)$$

where C_1 = coefficient from Table 20.34.

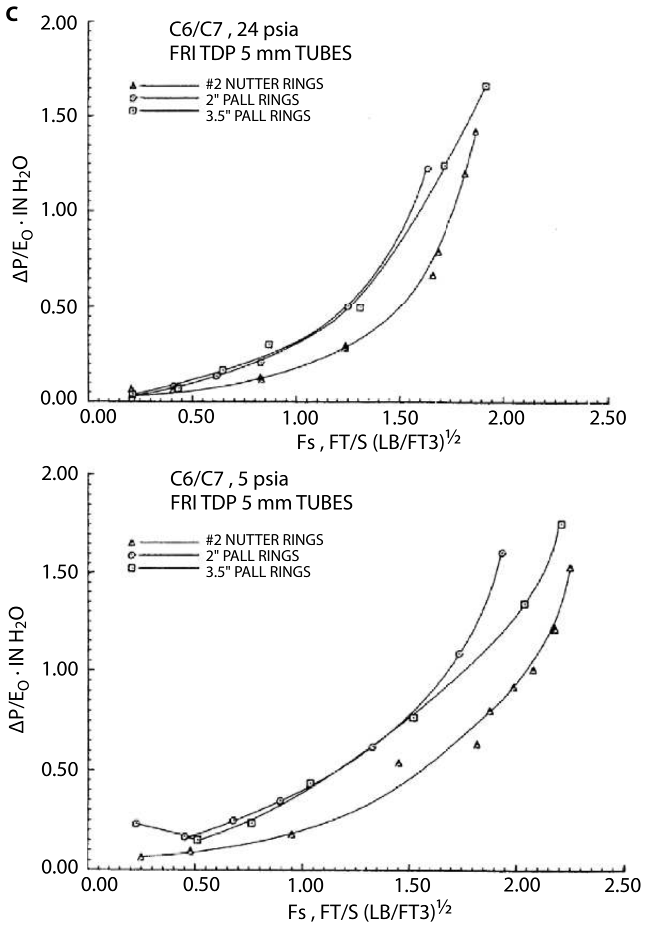


Figure 20.23C Comparison of P/E$_o$ for No. 2 Nutter Rings™ and Pall rings in a C_6/C_7 system at 24 psia and 5 psia using the FRI tubed drip pan distributor. Data prepared and used by permission of Nutter Engineering, Harsco Corp. and by special permission of Fractionation Research, Inc. All rights reserved.

Operating pressure drop:

$$\Delta P = e^{[(C_s - C_2)/C_3]}, \text{in. liquid/ft.} \tag{20.73}$$

where

C_3 = max. value of $[(Z_{Q1})\,(X_3)]$, or $[X_4$ itself$]$ (20.74)
C_3 = wet pressure drop slope coefficient
$Z_{Q1} = 0.1084 - 0.00350Q_1 + 0.0000438Q_1^2 + 7.67\times10^{-7}Q_1^3 - 1.4\times10^{-8}Q_1^4$ (20.75)
Z_{Q1} = liquid rate factor for C_3, in. water/ft
X_3, X_4 = constants from Table 20.34
ΔP_d = dry bed pressure drop, in. water/ft
ΔP = operating pressure drop, in. liquid/ft
e = base of natural logarithms
X_1, X_2 = curve fit coefficients for C_2, Table 20.34
X_3, X_4 = curve fit coefficients for C_3, Table 20.34
μ = viscosity, centipoises, cP

Subscripts
g,v = gas or vapor phase
l = liquid
s = based on tower cross-sectional area

Figures 20.24A and 20.24B [50] illustrate the correlation of wet pressure drop and system vapor rate at various liquid rates for No.2 Nutter rings; however, other available data indicate that other sizes of Nutter rings, Pall rings, and selected other packing shapes correlate in the same manner.

Figures 20.25A and 20.25B illustrate the fit of data taken by FRI on a commercial size column for hydrocarbon systems using No. 2.5 Nutter rings at three different pressures and comparing the latest Nutter proprietary correlation previously presented. When considering pressure drop models based only on water, hydrocarbons system capacity can be significantly overstated. For Nutter random ring packings, the pressure drop/capacity models fit the data within ±10% over the range of commercial interest, i.e., 0.1 to 1.0 in. water/ft of packing. Pressure drop values for design operation should not exceed 1.0 in. water/ft. This is generally 80% of flood capacity or 90% of useful capacity.

Tests by FRI and Nutter [52] emphasize that distribution of liquid must be uniform and at minimum values to achieve good HETP values over a range of system pressures for hydrocarbons distillation. Glitsch performance data for their Cascade Mini-Ring® (CMR) are shown in Figures 20.26A, B, and C for HETP with other published data and pressure drop for comparison with Pall rings and sieve trays.

Generally, it is not recommended to specify any packed section in a random packed tower to be greater than 20 ft in height. However, some packing manufacturers state that their packings will physically sustain greater heights and continue to produce good HETP values by maintaining a good uniform liquid flow internally, and that the liquid/vapor internal contacting does not create significant channeling that would reduce the contact efficiency. This statement is based on factors that include creating the distribution of liquid, redistribution of liquid, gas or vapor channeling, and process surging, plus many other situations unique to the process conditions. Structured packing heights should be determined by the manufacturer for the design conditions.

When trying to "balance" the several packed section heights required for the process, it is completely acceptable to vary the heights of each section to fit such requirements as location of return of reflux, multiple feed positions, and safety factors. Thus, one section may be 20 ft, another 17 ft, and another 25 ft as long as the process function has been thought through. The locations of the "breaks" in the packing sections must not interrupt control functions by

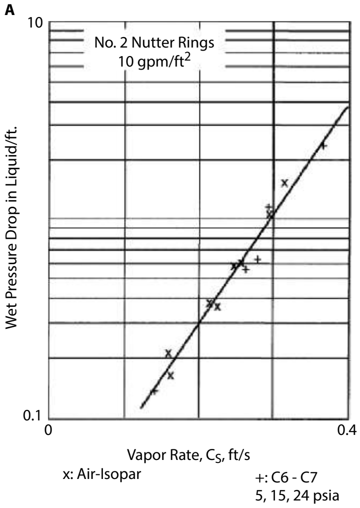


Figure 20.24A Correlation of No.2 Nutter Rings™ superficial capacity vs. wet pressure drop for four data sets and three separate tests. Note the 10:1 pressure drop range. Reproduced by permission from Nutter, D. E. and Perry, D., presented at New Orleans, La. meeting of American Institute of Chemical Engineers, March (1988), and by special permission of Fractionating Research, Inc. All rights reserved.

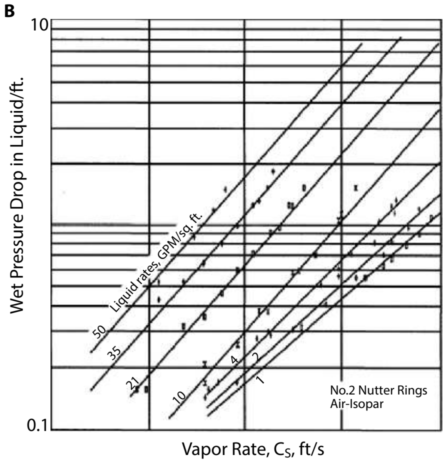


Figure 20.24B Correlation of No.2 Nutter Rings™ for 7 liquid rates versus wet pressure drop. Reproduced by permission from Nutter, D. E. and Perry, D., presented at New Orleans, La. meeting of American Institute of Chemical Engineers, March (1988), and by special permission of Fractionating Research, Inc. All rights reserved.

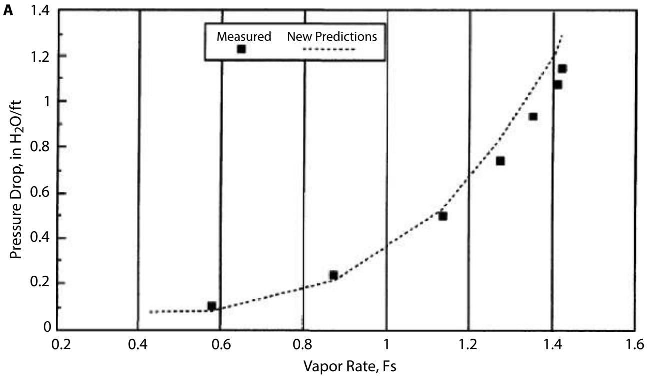


Figure 20.25A Comparison of hydrocarbon systems fit to Nutter Correlation at 165 psia (No. 2.5 Nutter Rings™). Used by permission of Nutter Engineering Co., Harsco Corp. and by special permission of Fractionation Research, Inc. All rights reserved.

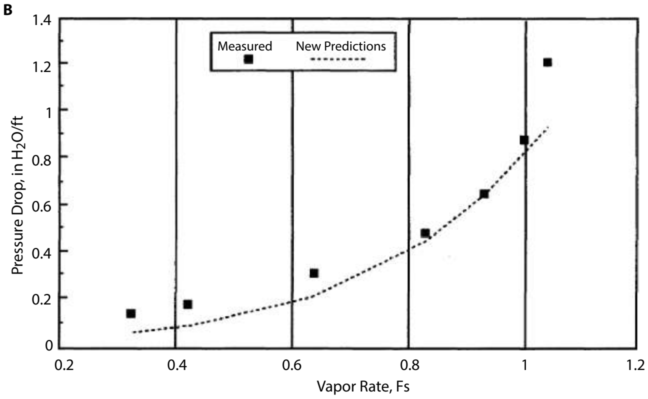


Figure 20.25B Comparison of hydrocarbon systems fit to Nutter Correlation at 300 psia (No. 2.5 Nutter Ring™). Used by permission Nutter Engineering Co., Harsco Corp. and by special permission of Fractionation Research, Inc. All rights reserved.

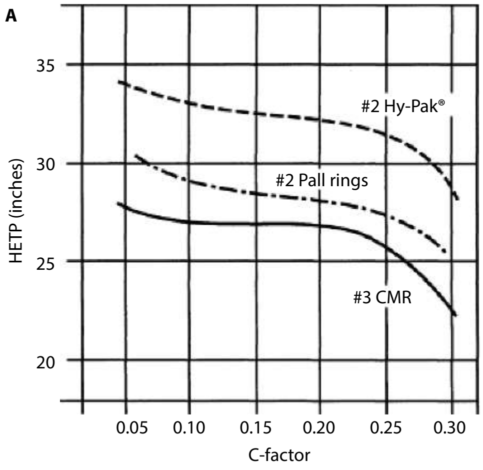


Figure 20.26A Efficiency versus C-factor for various metal packings. Iso-octane/toluene, 740 mm Hg, reflux ratio 14:1, 15-in., I.D. column, 10-ft bed height. Not Glitsch test data. Note; CMR = Cascade Miniring^{TM}. Used by permission of Glitsch, Inc., Bull. 345.

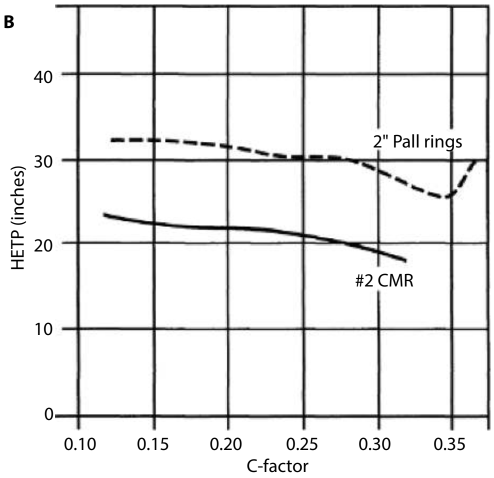


Figure 20.26B Efficiency versus C-factor for metal rings. Data from Billet (Symposium Series 32, Institution of Chemical Engineers), London (1969). Used by permission of Glitsch, Inc., Bull. 345.

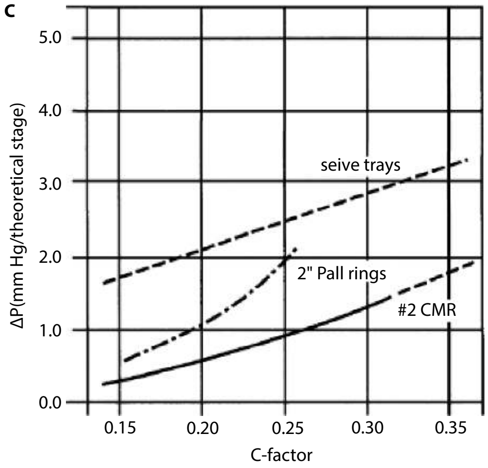


Figure 20.26C Pressure drop versus C-factor for metal rings and sieve trays. Operating data from ethylbenzene/xylene service. 50 mm Hg top pressure. Naarden International test data. Used by permission of Glitsch, Inc., Bull. 345.

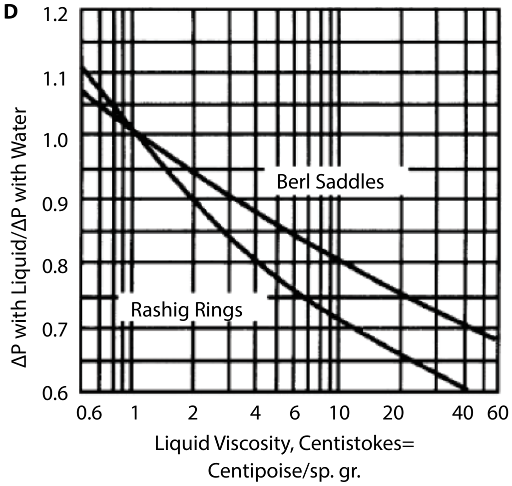


Figure 20.26D Pressure drop correlation at flood point for use with Table 20.33. Used by permission of Zenz, F. A., Chemical Engineering, Aug., (1953) p. 176. All rights reserved.

forcing the location of the temperature sensor. It is good to prepare a profile of the column and spot the controls and "interruptions" to packing continuity.

One issue is that at least three feed injection nozzles need to be placed in the column during packing (installation) and then piped up externally to the column's best performance determine conditions, and to allow for the monitoring of feed composition. In summary, everything that can be done mechanically including piping should be built into the column arrangement in order to be able to accommodate future changes in process requirements, errors, or lack of agreement between calculations and actual performance in operation.

20.22.2 Dumped Packing: Gas–Liquid System Below Loading

Figures 20.13B–H can be used effectively to determine performance for new designs or for existing operating columns.

20.22.3 Dumped Packing: Loading and Flooding Regions, General Design Correlations

Figure 20.21B, 20.13F, or 20.13H may be used for any system to obtain a good estimate of pressure drop for practically any random packing material. The relative liquid can position, and the approximate pressure drops per foot of packing depth may be read as parameters. It is important to recognize that the load upper limit, line A, is essentially coincident with the flooding condition. It is also apparent that the relative relation of the operating point to the flooding and loading conditions is quite different at the extreme right of the figure for large values of the abscissa than for low values to the left.

The rearranged form of the same Sherwood equation [53, 54] allows the curve for flooding of dumped packings to be presented so as to facilitate calculation of the flooding gas rate, G_f, corresponding to a given liquid flow L [55] (Figure 20.13A).

The packing factors to use with Figures 20.13B-F and 20.13H are given in Tables 20.26A–E.

Ward and Sommerfeld [56] present an equation based on the curves shown in Figures 20.13C and 20.13D and referenced to Eckert [9] and Leva [57] for calculating the gas and liquid flooding rates. There have been numerous other equations developed for this purpose, but many are too awkward for easy general use. The proposed equations have been tested by the authors.

Gas flow rate at flooding:

$$G = B\exp[-3.845186 + 4.044306(-0.4982244\ln_e(AB^2)-1)^{0.5}] \qquad (20.76)$$

liquid flow rate, L, at flooding:

$$L = \left[\frac{G}{(\rho_G/\rho_L)^{0.5}}\right]\exp[-4.303976 + 3.552134(-0.645854\ln_e(AG^2-1))^{0.5}] \tag{20.77}$$

where

A $= F\psi\mu_L^{0.2}/\rho_G\rho_L g_c$
B $= L(\rho_G/\rho_L)^{0.5}$
A = correlating parameter
B = correlating parameter
F = packing factor, 1/ft, or, $(ft)^{-1}$
G = gas flow rate, lb/(h) (ft^2)
g_c = Newton's Law, proportionality factor
= (4.17×10^8), $(ft\text{-}lb_m)/(lb_f\text{-}h^2)$
L = liquid flow rate, lb/(h-ft^2)
X = abscissa in the generalized pressure drop correlation
= $(L/G)(\rho_G/\rho_L)^{0.5}$
Y = ordinate in the generalized pressure drop correlation
$= G^2F\psi\mu_L^{0.2}/\rho_G\rho_L g_c$
μ = viscosity, centipoise
ρ = density, lb/ft^3
ψ = ratio of water density to liquid density

Subscripts
G = gas phase
L = liquid phase

The solution of the previous equations requires careful attention to the sequence of the arithmetic. One difficult requirement is the need to establish the L or G in 1b/h/ft^2 of tower cross-section, which requires a value for the tower diameter to be assured. The equations are quite sensitive to the values of A and B.

20.22.4 Dumped Packing: Pressure Drop at Flooding

As a comparison or alternative procedure, the pressure drop at the flooding point as indicated by the upper break in the pressure drop curve can be estimated from Table 20.36 and Figure 20.26D for rings and saddles [55]. The values in the table multiplied by the correction ratio give the pressure drop for the liquid in question, expressed as inches of water.

20.25.5 Dumped Packing: Pressure Drop Below and at Flood Point, Liquid Continuous Range

For a particular liquid–gas system and tower packing, performance indicates a region where the liquid phase becomes continuous and the gas phase discontinuous. This is obviously at relatively high liquid rates, but not beyond the range of satisfactory performance for the equipment. This region is characterized by proportionally higher-pressure drops than the gas-continuous region, and the existence of a critical liquid rate as this pressure drop deviation occurs.

Referring to Figure 20.12, the curve for L'-12,500 shows the beginning of the "move to the left," swinging away from the uniform slope of the curves for lower L' values.

This probably is not the L_c value for the system. The study of Zenz suggests that the critical liquid rate, L_c, is the minimum liquid rate that completely wets the packing, thus having essentially all packing surface available for gas contact. Rates above this value should be determined by allowable pressure drop and the limitation that the tower often begins to approach the flooding conditions more rapidly than in the gas-continuous region. Figure 20.27 correlates this L_c for Raschig rings and Berl saddles as a function of liquid viscosity.

More work is needed to fully understand this feature of tower performance and to extend the existing information to other common packings. Determination of L_c from the figures will indicate whether the tower is operating under gas-continuous (values of L lower than L_c) or liquid-continuous (values equal to or larger than L_c). The approximate degree of wetting of the packing can be evaluated as the ratio of L/L_c [58]. The pressure drop is evaluated using Figures 20.13B–20.13F or 20.13H to determine the flooding liquid rate, L_f. Then calculate the ratio of L_c to actual L. Read Figure 20.28 to obtain ΔP actual/ΔP_f. Thus, ΔP actual is the ratio value times ΔP_f calculated using Figure 20.26D and Tables 20.23, 20.23A, and 20.35.

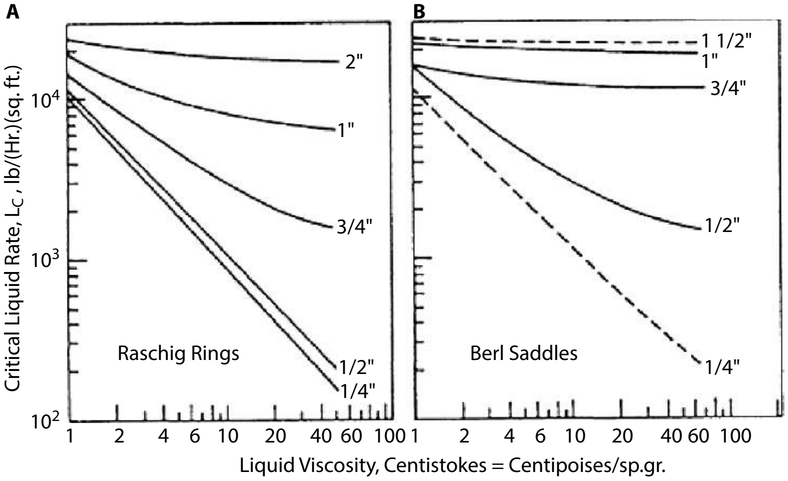


Figure 20.27 Values of liquid rate when the system becomes liquid continuous, Lc. Used by permission of Zenz, F. A., Chemical Engineering, Aug (1953) p. 176. All rights reserved.

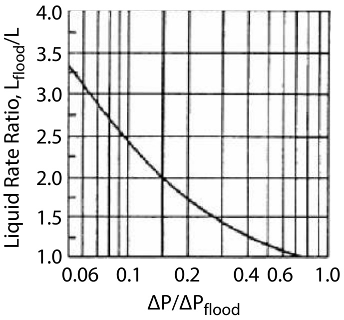


Figure 20.28 Pressure drop correlation at flooding. Used by permission of Zenz, F. A., Chemical Engineering, Aug (1953) p. 181.

Table 20.35 Pressure drop at upper break point (flood) with water as the flowing liquid.

Raschig rings, in.	ΔP_f, in. H_2O/ft of packed bed	Berl saddles, in.	ΔP_f, in. H_2O/ft of packed bed
2	2.5	1 ½	2.2
1 ½	2.5	1	2.5
1 ¼	4.0	¾	2.5
1 (ribbed)	3.0	½	2.0
1	4.0	¼	1.25
¾	3.0		
5/8	2.5		
½	3.5		
3/8	4.0		
¼	4.0		

By permission, F. A. Zenz, *Chem. Eng*., Aug., 176 (1953), Ref. 81.

20.22.6 Pressure Drop Across Packing Supports and Redistribution Plates

Little useful correlated information on pressure drop across packing supports and redistribution plates is available. Some order of magnitude guide data are given in Figures 20.29 to 20.33.

Because these data are peculiar to the supports studied, they can serve only as a good estimate for other situations. It is important to remember that the results obtained with a bare plate, and one holding a layer of packing, can be quite different, the latter being the more realistic condition.

The only available test data [59] indicate that the plain flat plate (26-45% free area) has a decidedly detrimental effect on the allowable flooding conditions of the tower, whereas the wire screen, weir-type, and fused Raschig ring designs have very little effect when using Intalox saddles for packing.

In general, the dumped saddle type packing should give less blocking to support openings than ring type.

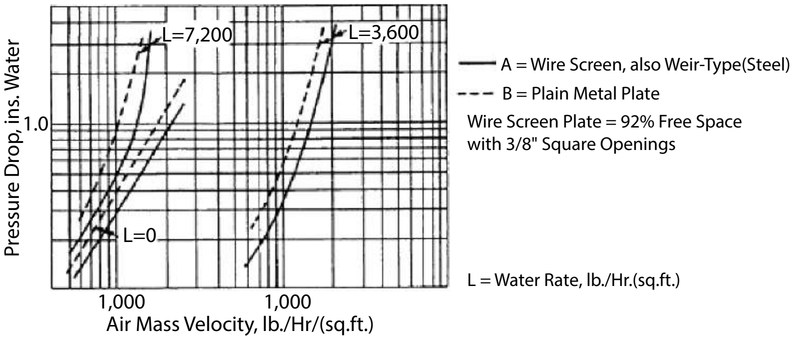


Figure 20.29 Comparison effect of pressure drop across support plate and bed of 1 ½ in. Intalox® saddles. Used by permission of Leva, M., Lucas, J. M., and Frahme, H. H., Ind. Eng. Chem. V. 46, No. 6 (1954). All rights reserved.

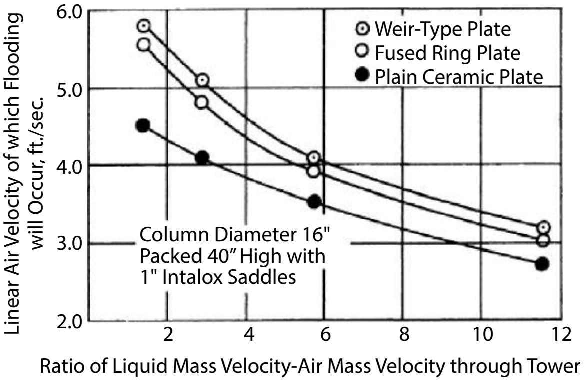
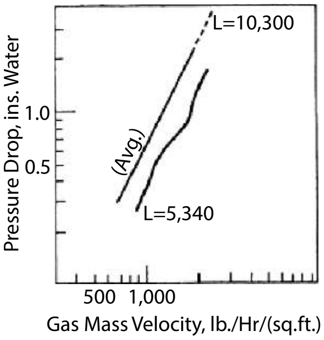


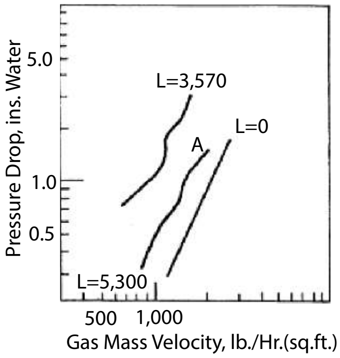
Figure 20.30 Effect of choice of support plate on flooding rate. Used by permission of U.S. Stoneware Co. (now Norton Chemical Process Equipment Corp.).



Pressure Drop Through Fused Raschig
Ring Plate, with Plate Covered with
One inch of 1" Size Intalox Saddles
(L=Liquid Mass Rate Parameter)

Plate = 58% Free Area

Figure 20.31 Pressure drop through fused ceramic Raschig ring plate with plate covered with 1 in. of 1-in. Intalox® saddles. Used by permission of Leva, M. Lucas, J. M. and Frahme, H. H., Ind. Eng. Chem., V. 46, No. 6 (1954); all rights reserved.



Pressure Drop through Plain Ceramic
Plate
(L=Liquid Mass Rate Parameter)
Curve A also Represents Weir-Type
Plate Covered With One inch of 1"
Intalox Saddles
Plate = 20% Free Area Plain Ceramic
= 60% Free Area Weir-Type

Figure 20.32 Pressure drop through plain ceramic plate. Used by permission of Leva, M., Lucas, J. M., and Frahme, H. H., Ind. Eng., Chem. V. 46, No. 6 (1954). All rights reserved.

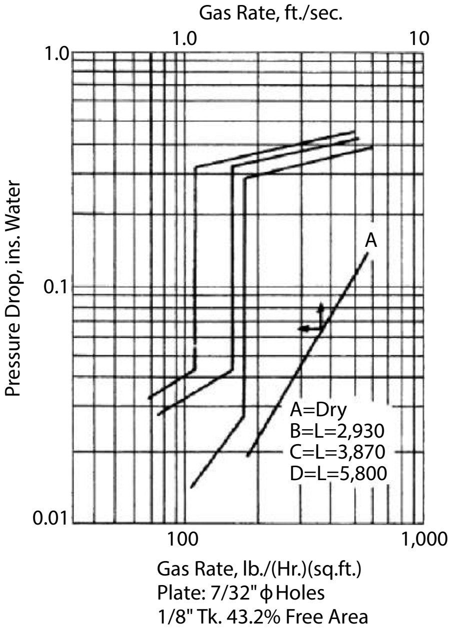


Figure 20.33 Flooding on perforated support plate. Used by permission of Lerner, B. J. and Grove, C. S. Jr., Ind. Eng. Chem. V. 43, No. 1 (1951), p. 216. All rights reserved.

Example 20.3: Evaluation of Tower Condition and Pressure Drop

Check the design of a 4-ft, 6-in. I.D. tower packed with 45 ft of 1-in. x ⅙-in. thick steel Raschig rings if the service requires a liquid rate of 2,250 lb/h (ft^2) of 10% caustic solution (sp. gr. = 1.22) and 4,540 lb/h (ft^2) at 110°F air containing CO_2 to be scrubbed at 365 psia.

1. Determine the operating range for the tower by referencing to Figure 20.13B or 20.13C. Use Figure 20.13C for this example.

Solution:

$$\rho_G = \left(\frac{29}{379}\right)\left(\frac{520}{570}\right)\left(\frac{365}{14.7}\right) = 1{,}732\ \text{lb/ft}^3$$

$$\rho_L = (62.3)\ (1.22) = 76.1\ \text{lb/ft}^3$$

$$\mu = 2\ \text{cP at } 104^\circ\text{F for liquid}$$

$$F = \frac{a}{\varepsilon^3} = 137\ \text{ Table 20.25A note range.}$$

$$\frac{L}{G}[\rho_G/(\rho_L - \rho_G)]^{1/2} = \frac{2250}{4540}\left(\frac{1.732}{76.1 - 1.732}\right)^{1/2} = 0.07563$$

$$\frac{G_2}{\rho_G}\left(\frac{a}{\varepsilon^3}\right)\frac{\mu^{0.1}}{(\rho_L-\rho_G)g_c}=\frac{(4540/3600)^2-(1.07)^{0.1}(137)}{1.732(76.1-1.732)(32.2)}=0.0563$$

2. Locating 0.0756 and 0.0563 on Figure 20.13C reading the intersection indicates a condition in the lower loading region, and a pressure drop of approximately 0.60 in. H_2O/ft of packing.
3. Expected pressure drop through bed: = (45 ft) (0.60) = 27 in. H_2O, total
4. This bed should be in three sections (two might be acceptable under some circumstances, or if different packing were used), thereby requiring two intermediate combined packing supports and redistribution plates and one bottom support plate.
   Estimate pressure drop per redistribution set-up or support = 1.0 in.
   For two redistributors plus one support plate:
   Total estimated pressure drop = (3) (1.0) = 3 in. H_2O
5. Total estimated pressure drop through packed portion of tower:

   Bed ΔP = 27.0
   Internals = 3.0
   30.0 in. H_2O

   Say, 30 to 35 in. H_2O (this should be 30.0 ±15-20%)
6. Estimated percent of flooding
   Reading the abscissa of Figure 20.13C at 0.0756 for ordinate at flooding line for dumped packing: = 0.13
   Then, actual value of 0.0563 is
   Percent of liquid flooding = $\frac{0.0563}{0.13}(100)=43.3\%$
   Note that this rather low value will usually occur when operating at the low L/G values due to the greater operating spread between flooding and loading (see Figure 20.13C).
7. Estimated percent of loading (average)
   Reading abscissa at 0.0756 for ordinate of line B gives ordinate = 0.078
   Percent of flooding $=\frac{0.0563}{0.078}(100)=72.1\%$ (not very precise)
   This is an acceptable value and should not be exceeded except in known systems. It is preferable to operate reasonably close to the loading condition for maximum best efficiency of contact.
   As an alternate consideration, assume various pressure drops/foot of packing (same) and determine the effect on calculated column diameter. Use the same input information as original stated conditions. Then:

Assumed Pressure Drop, in. H_2O/ft	**Calculated Column Diameter, ft**
0.25	5.22
0.50	4.61
0.75	4.35

Example 20.4: Alternate Evaluation of Tower Condition and Pressure Drop

Using Example 20.3 as reference, the tower will be examined for critical liquid rate, L_c, using Figure 20.27.

$$\text{Centistokes}=\left(\frac{\text{centipoise}}{\text{sp gr}}\right)\text{ for caustic solution.}=\frac{2}{1.22}=1.64$$

reading: L_c = 16,000 lb/(h) (ft^2)

Because actual L = 2250 is less than 16,000, this tower operates in the gas-continuous region.

Pressure drop at flooding from Table 20.35 and Figure 20.26D.

$\Delta P_f = 4$ in. H_2O/ft (not exact figure, because table is for ceramic rings)

Correction = 0.94

Then actual expected pressure drop at flooding:

$$\Delta P_f = (0.94)\ (4) = 3.76 \text{ in. } H_2O$$

For 45 ft of packing

$$\Delta P_f \text{ total} = (45)\ (3.76) = 169 \text{ in. } H_2O$$

Comparison with Figure 20.13C gives 3 in. H_2O/ft (parameter) or a total of (3) (45) = 135 in. H_2O. Neither of these values represents a condition (flooding) that should be considered for tower operation, except for very well-known systems. Distillation operations sometimes operate above flooding, but other types of contacting normally require operations in the loading region (or below) for stable performance.

Example 20.5: Change of Performance with Change in Packing in Existing Tower

A tower is packed with l-in. ceramic Raschig rings. It presently floods while drying product at a production feed rate of 1,800,000 lb/month with 0.25 mol% being water. Flooding does not start at the bottom, but at some intermediate point up the tower. What can be done to eliminate the flooding? Is it possible to increase production rate to 2,000,000 lb/month?

Solution

1. Examine packing characteristics.

Size	**a/ε³**	**Surface area, ft²/ft³**
1 in. Rashig Rings	158	58
1 in. Intalox	124	78
1 in. Berl Saddles	125	76
1 in. Pall Ring*	45	66.3
1½ in. Intalox	69	59.5
1½ in. Berl Saddle	79	44

*metal, all others ceramic.

2. Percent of flooding for various packings, holding tower flow rates (including reflux) constant. Refer to Flooding, Loading, and Pressure Drop Chart, Figure 20.13D.

$\frac{L}{G}(\rho_G/\rho_L)^{0.5}$ remains constant for same separation at increased production rate.

$\frac{G^2\psi^2\mu^{0.2}}{\rho_G\rho_{Lgc}}\left(\frac{a}{\varepsilon^3}\right)$ increases as G^2 at increased production rate for a fixed a/ε^3.

Packing	**Percent of Flooding at 2,000,000 lb/mo. rate referred to flooding at 1 in. ceramic Raschig rings**
1 in. R.R.	= 100%
1 in. Intalox	$\left(\frac{2}{1.8}\right)^2\left(\frac{124}{158}\right)(100) = 96.9\%$
1 in. Pall Rings (metal)	$\left(\frac{2}{1.8}\right)^2\left(\frac{45}{158}\right)(100) = 35.2\%$
1½ in. Intalox	$\left(\frac{2}{1.8}\right)^2\left(\frac{69}{158}\right)(100) = 53.9\%$
1½ in. Berl Saddle	$\left(\frac{2}{1.8}\right)^2\left(\frac{79}{158}\right)(100) = 61.8\%$

The flooding of the packing is a direct function of the a/ε^3; therefore, it is valid at constant separation to examine the performance as shown. The metal Pall rings appear to allow for a considerable increase in capacity. In fact, the condition at 35.2% of flood might not be good from a contact efficiency standpoint.

3. Selection

The 1½-in. Intalox or Berl ceramic saddles would be the preferred choice because: (1) the flooding point is sufficiently low and yet probably not too far from the load point (only flood data available, but would estimate 70–85% of load); (2) the surface area per cubic foot is essentially the same as for the existing 1-inch Raschig rings. By reference to the effective interfacial area graphs, and by using the Berl saddle data instead of Intalox as an estimate because it is not available, the separation performance would be expected to be essentially identical to the existing tower (for the lntalox); (3) the production rate can be increased; (4) the flooding can be removed from present operations using the 1½-in. Intalox or any of the other packings, but note that the a/ε^3 of the Berl saddles is not as high as the l-in. R. R. This might mean less efficient contact, requiring more packing. However, from Figure 20.34, note that for a given L′/G′, the Berl saddles are from 15 to 30% better wetted in any given packing volume. From Figures 20.35 and 20.36, the comparative effective areas indicate that the Berl saddles have about the same area/ft^3 as the existing rings. Therefore, it appears that the flooding can be stopped (lower a/ε^3) and yet the contact can be as good, or may be a little better than it was originally.

4. Check support grid voidage

The packing support consists of a floor grating material with bar openings spaced to give 57.5% free void area of cross-section.

To avoid the possibility of local flooding at the support, it would be well to place a heavy hardware cloth over this grid to keep the saddles from nesting at the first layer in the opening slots. It is preferable to have them resting on a surface that cannot be easily blanked.

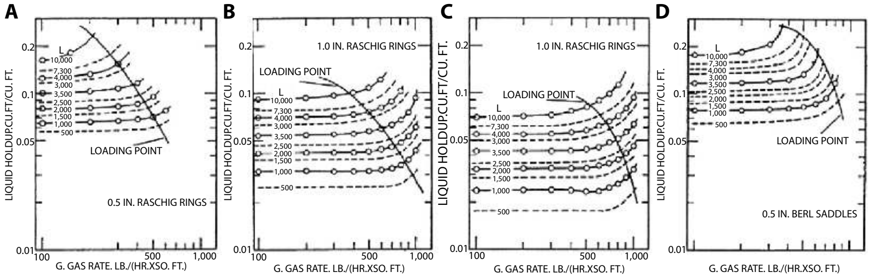
Select a l-in. x l-in. (center line) x 0.063-in. wire cloth with voidage of 87.8%.

Combined voidage, support grid plus cloth:

$$= (0.575)\ (0.878) = 50.5\%$$

This should be satisfactory, but a value much lower than this could not be tolerated.

Dump Intalox (or Berl) saddles into tower while tower is filled with water.

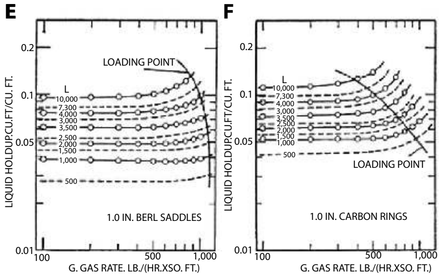


Total holdup in 0.5 inch Raschig rings.

Total holdup in 1.0 inch Raschig rings.

Total holdup in 1.5 inch Raschig rings.

Total holdup in 0.5 inch Berl saddles



Total holdup in 1.0 inch Berl saddles

Total holdup in 1.5 inch carbon rings

Figure 20.34 Total hold-up for ceramic tower packings: air-water. Used by permission of Shulman, H. L., Ullrich, C. F., and H. Wells., A.IChE, Jour., V. 1, No. 2. (1955) p. 247. All rights reserved.

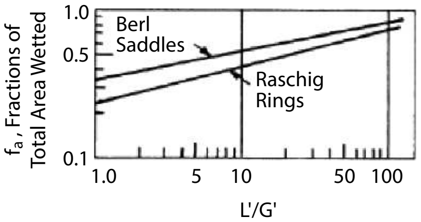


Figure 20.35 Fraction packing wetted. Used by permission of Shulman, H. L., Ullrich, C. F., Proulx, A. Z. and J. O. Zimmerman, A.IChE, Jour., V. 1, No. 2. (1955) p. 253. All rights reserved.

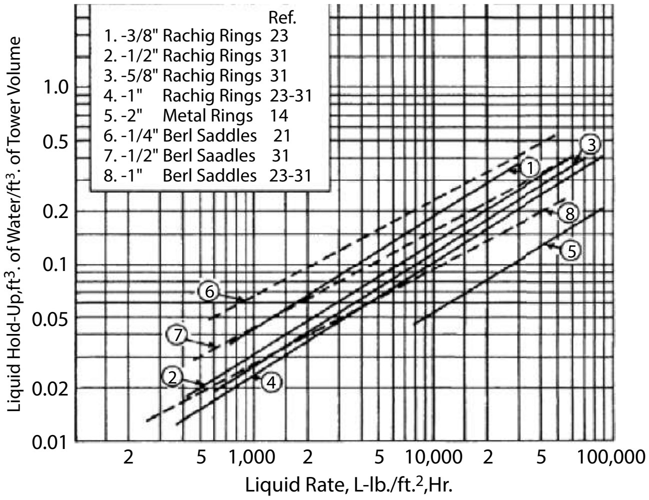


Figure 20.36 Gas–liquid hold-up data for ceramic rings and saddles. Used by permission of Leva, M. Tower Packings and Packed Tower Design, 2nd. Ed., U.S. Stoneware Co. (now, Norton Chemical Process Equipment Corp.) 1953.

Example 20.6: Stacked Packing Pressure Drop

Consider the problem of drying air with sulfuric acid (Example 20.7). If the mass transfer relations are evaluated, a reasonably good estimate of the new packed height can be determined for using 2-in. stacked ceramic Raschig rings in place of Intalox saddles. For the present, assume the new required stacked height is 25 ft. Although the pressure drop per foot will be less and the flooding point higher than the dumped packing, it is inadvisable to go to a smaller diameter tower because of the high superficial gas velocity.

Using the 15-in. I.D. ceramic tower, the expected pressure drop will be

$$\Delta P = (\alpha)(10^{\beta L})\left(\frac{G^2}{\rho_G}\right) \tag{20.12}$$

From Reference 31 for 2-in. ceramic stacked Raschig rings

$$\alpha = 0.06$$

$$\beta = 0.012$$

Liquid range checks satisfactory

$$G = 0.593 \text{ lb/(s) (ft}^2)$$

Leva [31] has shown that for liquids other than water, the L must be corrected by the ratio of the density of water to that of the fluid in the system.

$$L = 0.378 \text{ lb/(s)(ft}^2)\left(\frac{62.3}{112.6}\right) = 0.209$$

$$\rho_G = 0.087 \text{ lb/ft}^3$$

$$\Delta P = (0.06)(10^{(0.12)(0.209)})\left(\frac{(0.593)^2}{0.087}\right)$$

$$\Delta P = 0.00258 \text{ in. water/ft packed height (estimated)}$$

Total tower drop:

Packing	= (0.00258) (25) = 0.064 in. H_2O (*approximate*)
Support (estimated)	= 1.5 in. H_2O
Total (approximate)	1.56 in. H_2O

Note that the weight of liquid will be greater in this arrangement at flooding, and the operating hold-up will be almost the same as the dumped Intalox. The total weight of packing will be approximately 50% greater than if the same 2-in. rings had been dumped in place. Two-inch rings are not usually stacked. In this small tower made up of 3-ft ceramic sections, the stacking is not too difficult a job if there are conditions that justify the extra effort and expense.

20.23 Liquid Hold-Up

Liquid hold-up in a tower represents the liquid held in the void spaces of the packing during operating conditions. At flooding, essentially all of the voids are filled with liquid.

Usually low hold-up is desired but reasonable hold-up is necessary for efficient tower operation. The weight of liquid held in the packing must be considered when determining the support loads at the bottom of the packing, as well as the tower itself. The higher the hold-up for any particular packing, the greater will be the gas pressure drop, and the longer the tower drainage time when shut down. Smaller size packing tends to have greater hold-up than larger packing.

Figure 20.36 presents water hold-up data that are correlated by [31]

$$h_{tw} = 0.0004\left(\frac{L'}{d_p}\right)^{0.6} \tag{20.78}$$

where

h_{tw} = water hold-up, ft^3 liquid/ft^3 volume tower
L' = Liquid rate, lb/(h) (ft^2)
d_p = Equivalent spherical packing diameter, inches (diameter of packing equivalent to sphere of same surface area as the packing piece). Values given for some packings with physical data Tables 20.1 through 20.14. Area of sphere = π (diameter)^2.

For liquids other than water [31, 60]:

$$h_1 = h_{tw}(\mu_L)^{0.1}\left(\frac{62.3}{\rho_L}\right)^{0.78}\left(\frac{73}{\sigma}\right)^{n} \tag{20.79}$$

where

h_l = 1iquid hold-up, ft^3/ft^3 packed tower volume
μ_L = liquid viscosity, centipoise
ρ_L = liquid density, lb/ft^3
σ = surface tension, dynes/cm

Values of exponent n are given in Figure 20.37.

Total liquid hold-up in packed bed, h_t = static hold-up, h_s, plus operating hold-up, h_o [61, 62].

The static hold-up is independent of liquid and gas rates, since it represents the liquid held in the packing after a period of drainage time, usually until constant weight of material is received. This requires approximately 1 hour for a 10-in. dia. x 36-in. packed height tower. Table 20.36 adequately summarizes the data.

Total hold-up, h_t, of water is represented for Raschig rings and Berl saddles [62].

$$h_t = \alpha L'\beta/D_p^2 \tag{20.80}$$

$$\beta = \gamma D_p\theta \tag{20.81}$$

Constants are given in Table 20.37.

Figures 20.34A–F present the graphical interpretation of the total hold-up equation. These are more convenient to apply where the system fits (or nearly fits) the curves.

These data are valuable for determining the total weight of liquid held in the packing, and also the void fraction, in an operating column. ε is the void fraction of the dry packing minus the total hold-up, h_t.

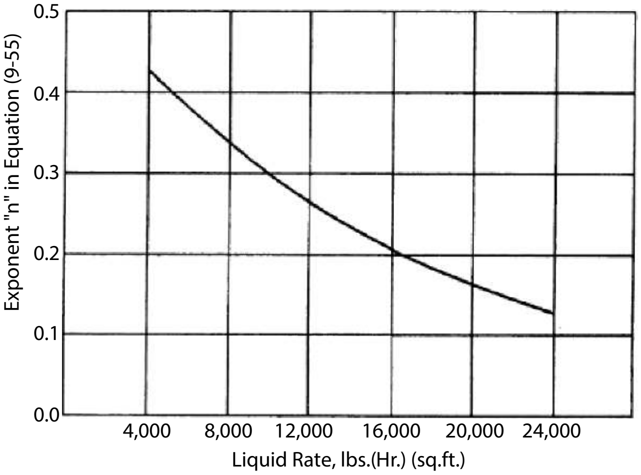


Figure 20.37 Liquid hold-up variation of surface tension exponent with liquid rate. Reproduced by permission of the American Institute of Chemical Engineers; Jesser, B. W. and Elgin, J. C., Trans. A.I.ChE., V. 39, No. 3 (1943) p. 295. All rights reserved.

Table 20.36 Static hold-up in random packings.

Packing nominal size, in.	**Static water hold-up, h_s, ft^3/ft^3 packing**
Raschig rings (unglazed porcelain)	
½	0.0325
1	0.0150
1 ½	0.0089
2	0.0058
Berl saddles (unglazed porcelain)	
½	0.0317
1	0.0110
1 ½	0.0052
Raschig rings (carbon)	
1	0.0358
1 ½	0.0200
2	0.0150

By permission of The American Institute of Chemical Engineers, Shul-man, H. L. *et al.*, *Chem. Engr. Jour*. Vol. 1, No. 2, 247 (1955) and ibid, p. 259 (1955), Ref. 64 and 66, all rights reserved.

Table 20.37 Total hold-up constants.

Packing	**α**	**ϒ**	**θ**
Porcelain Raschig ring	2.25×10^{-5}	0.965	0.376
Carbon Raschig ring	7.90×10^{-5}	0.706	0.376
Porcelain Berl saddles	2.50×10^{-5}	0.965	0.376
Porcelain, in.	**Equivalent spherical dia., ft**		
½ R. R.	0.0582		
1 R. R.	0.1167		
1 ½ R. R.	0.1740		
2 R. R.	0.238		
½ Berl saddle	0.0532		
1 Berl saddle	0.1050		
1 ½ Berl saddle	0.155		
Carbon, in.			
1 R. R.	0.1167		
1 ½ R. R.	0.178		
2 R. R.	0.235		

By permission of The American Institute of Chemical Engineers, Shulman, H. L. et al., *Chem. Engr. Jour*. Vol. 1, No. 2, 247 (1955) and ibid, p. 259 (1955), Ref. 64 and 66, all rights reserved.

20.23.1 Correction Factors for Liquids Other Than Water

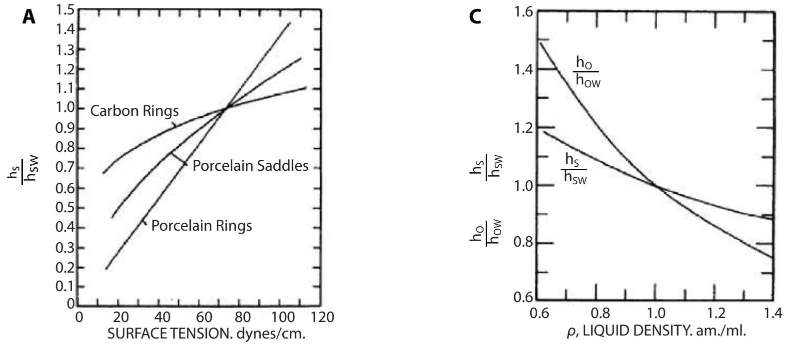
In order to use the data in systems handling liquids other than water, correction equations and charts are used [62]. The charts are more convenient to use and are presented in Figures 20.38A–D. First, determine the total or static hold-ups for water at 20°C; second, determine separately the correction for viscosity, density, and surface tension; third, multiply the water hold-up by each of the corrections to obtain hold-up for liquid of the specific system.

A second hold-up correlation reported by T. Otake and K. Okada [63] represents a survey of considerable literature and is applicable to aqueous and non-aqueous systems for Reynolds numbers from 10, -20,000 [31].

$$h_o = 1.295\left(\frac{\left(D'_pL\right)^3}{\mu'}\right)^{0.676}\left[\frac{\left(D'_p\right)^3 g_c\rho^2}{\mu'^2}\right]^{-0.44}\left(aD'_p\right) \tag{20.82}$$

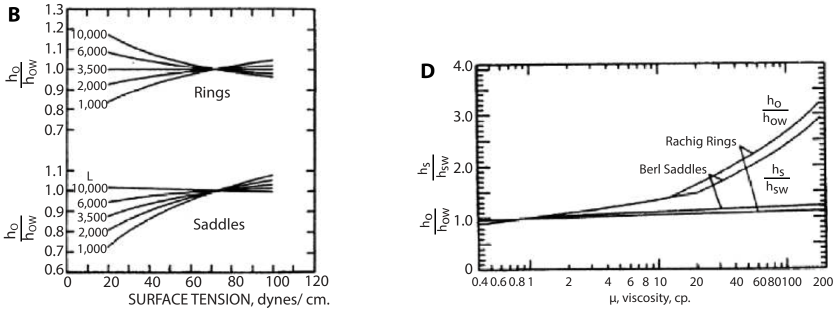
20.24 Packing Wetted Area

Wetted packing area may be considerably different to the physical area of a packing. This is of particular importance in comparing the effectiveness of different packings. It is only recently that a coordinated group of data became available for wetted areas in Raschig ring and Berl saddle packing [64].



A. Factors to be applied to water static hold-ups to determine the effect of surface tension. Used by permission of American Institute of Chemical Engineers, *A.I.Ch.E Jour.*, Shulman, H. L., Wells, N., Ullrich, C. F., and Proulx, A. Z., V. 1, No. 2 (1955) p. 259; ALL right reserved.

C. Factors to be applied to water hold-ups to determine effect of liquid density. Used by permission of American Institute of Chemical Engineers, *A.I.Ch.E Jour.*, Shulman, H. L., Wells, N., Ullrich, C. F., and Proulx, A. Z., V. 1, No. 2 (1955) p. 259; all right reserved.



B. Factors to be applied to water operating hold-ups to determine effect of surface tension. Reproduced by permission of the American Institute of Chemical Engeineers, *A.I.Ch.E Jour.*, Shulman, H. L., Wells, N., Ullrich, C. F., and Proulx, A. Z., V. 1, No. 2 (1955) p. 259; ALL right reserved.

D. Factors to be applied to water hold-ups to estimate effect of liquid density. Used by permission of American Institute of Chemical Engineers, *A.I.Ch.E Jour.*, Shulman, H. L., Wells, N., Ullrich, C. F., and Proulx, A. Z., V. 1, No. 2 (1955) p. 259; all right reserved.

Figure 20.38 Physical property corrections for liquid hold-up for ceramic packing and carbon packing (as noted). Reproduced by permission of A.I.Ch.E. Jour., Shulman, H. L., Wells, N., Ullrich, C. F., and Proulx, A. Z., V. 1, No. 2 (1955) p 259. All rights reseved.

Figure 20.35 represents the water-air and ammonia–water data for Berl saddles by [64]

$$f_a = \frac{a'}{a_t} = 0.35\left[\frac{L'}{G'}\right]^{0.20} \tag{20.83}$$

and for Raschig rings by

$$f_a = \frac{a'}{a_t} = 0.24\left[\frac{L'}{G'}\right]^{0.25} \tag{20.84}$$

where

f_a = fraction of total packing area, a_t, that is wetted
a' = wetted packing surface not necessarily same as effective interfacial surface area for contact, ft^2/ft^3
a_t = total packing surface, ft^2/ft^3
L' = superficial liquid rate, lb/h (ft^2)
G' = superficial gas rate, lb/h (ft^2)

The fraction wetted area immediately indicates the effectiveness of contact for the liquid system in the packing. This contact efficiency must be considered in some design problems.

20.25 Effective Interfacial Area

The effective interfacial area is used in mass transfer studies as an undivided part of individual and overall coefficients when it is difficult to separate and determine the effective area. The work of Shulman *et al.* [64] presents a well-organized evaluation of other work in addition to their own. One of the difficulties in correlating tower packing performance lies in obtaining the correct values for the effective interfacial areas of the packing on which the actual absorption, desorption, chemical reaction, etc. are completed. Figures 20.39A–G present a correlation for water flow based on the ammonia–water data of Fellinger [120] and are valid for absorption work.

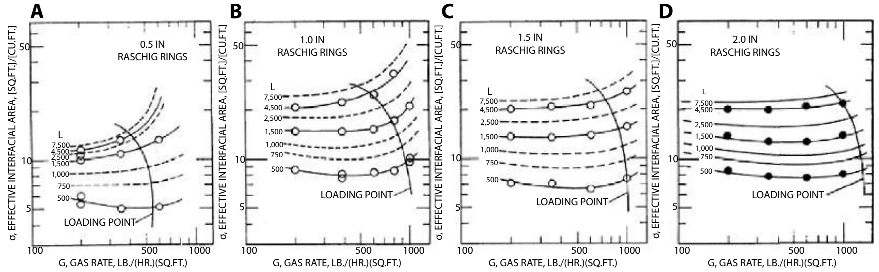
There are differences between wetted and effective area as discussed by Shulman [64]:

1. Wetted areas increase as packing size decreases.
2. Effective area is smallest for the smallest packings.
3. The effective area seems to go through a maximum for the l-inch size packing, although the larger packings have almost as much area.

This is better understood in terms of the hold-up data for these packings. For vaporization in packed beds of Raschig rings and Berl saddles [62]:

$$a_{vap} = 0.85(a_{abs})\left(\frac{h_t}{h_o}\right) \tag{20.85}$$

$$a_{vap} = 0.85(a_{abs})_w\left(\frac{h_t}{h_{ow}}\right) \tag{20.86}$$

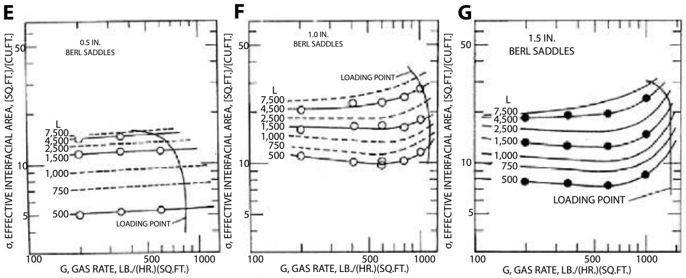


Effective interfacial area for 0.5 in. Raschig rings.

Effective interfacial area for 1-in. Raschig rings.

Effective interfacial area for 1.5 in. Raschig rings.

Effective interfacial area for 2.0-in. Raschig rings, using estimated hold-ups.



Effective interfacial area for 0.5 in. Berl saddles.

Effective interfacial area for 1.0-in. Berl saddles.

Effective interfacial area for 1.5-in. Berl saddles, using estimated hold-ups.

Figure 20.39 Effective interfacial areas for random ceramic tower packings. Note for gases or vapors other than air, use abscissa, G, as $G/(\rho_{gas}/0.075)^{0.8}$. Based on the data of Fellinger [27]. Used by permission of the American Institute of Chemical Engineers; A.I.ChE., Jour. Shulman, H. L., Ullrich, C. F., Proulx, A. Z, and Zimmerman, J. O., V. 1, No. 2, (1955) p253. All rights reserved.

where the subscripts *vap* and *abs* represent conditions of vaporization and absorption, respectively, and subscript w represents a water system.

20.26 Entrainment from Packing Surface

There is not much data available for this subject; operational experience plus qualitative tests indicate that entrainment is negligible until the packing reaches the flooding condition. See discussion under distillation section.

Example 20.7: Operation at Low Rate, Liquid Hold-Up

A sulfuric acid drying tower uses 98% acid for drying an incoming air stream. The pilot plant tests show that 15 ft of l-in. ceramic Intalox packing will do this job. The plant scale rates are

Air = 500 scfm at 90°F and 2 psig
Acid = 6 gpm at 90°F and sp gr = 1.81

Determine (1) the tower diameter, (2) pressure drop, and (3) liquid hold-up.

Solution

$$\text{air rate} = \frac{500}{359}\left(\frac{29}{60}\right)\left(\frac{460+60}{460+90}\right)\left(\frac{14.7+2}{14.7}\right) = 0.725\ \text{lb/s}$$

$$\text{liquid rate} = \left(\frac{18}{60}\right)\frac{(8.33)}{1.81} = 0.461\ \text{lb/s}$$

Assume for first attempt:

$$\text{Inside tower diameter} = 12\ \text{in.}$$

$$\text{Cross-section area} = \frac{\pi(1)^2}{4} = 0.785\ \text{ft}^2$$

$$\text{Then, air rate, G} = \frac{0.725}{0.785} = 0.925\ \text{lb/(s)(ft}^2\text{)}$$

$$= 3340\ \text{lb/(h) (ft}^2\text{)}$$

$$\text{liquid rate, L} = \frac{0.461}{0.785} = 0.588\ \text{lb/(s)(ft}^2\text{)}$$

$$= 2120\ \text{lb/(h) (ft}^2\text{)}$$

$$\rho_L = (1.81)\ (62.3) = 112.6\ \text{lb/ft}^3$$

$$\rho_G = \frac{29}{359}\left(\frac{460+60}{460+90}\right)\left(\frac{14.7+2}{14.7}\right) = 0.087\ \text{lb/ft}^3$$

$$\frac{L}{G}(\rho_G/\rho_L)^{1/2} = \frac{0.588}{0.925}\left(\frac{0.087}{112.6}\right)^{1/2} = 0.01975, (\text{abscissa})$$

$$\left(\frac{a}{\varepsilon^3}\right) \text{for 1-in. Intalox saddles} = 92 \text{ (Table 20.26A)(original value)}$$

$$\psi^2 = \left(\frac{1}{1.81}\right)^2 = 0.305$$

$$g_c = 32.2\ (\text{lb}_m)\ (\text{ft})/(\text{lb}_f)\ (\text{s})^2$$

$$\mu^{0.2} = (17.0)^{0.2} = 1.762 \text{ cP at } 90°\text{F}$$

$$\frac{G^2}{\rho_G}\left(\frac{a}{\varepsilon^3}\right)\frac{\psi^2\mu^{0.2}}{\rho_L g_c} = \frac{(0.925)^2(92)(0.305)(1.762)}{(0.087)(112.6)(32.2)} = 0.13411(\text{ordinate})$$

Reading Figure 20.13D at the calculated ordinate and abscissa, we obtain:

(a) An indicated operating condition above the upper loading limit
(b) A condition of high-pressure drop, approximately 1.5 in./ft of packing height
(c) A situation too close to flooding, thus requiring a larger tower diameter

Second Attempt: Try 15-inch Dia. Ceramic Tower

Inspection of Figure 20.13B, C, or D shows that the increase in tower diameter is not affected by the value of the abscissa. By changing the tower diameter to 15-in. cross-section area = 1.22 ft^2.

$$G = \frac{0.725}{1.22} = 0.593 \text{ lb/(s)(ft}^2)$$

$$L = 0.461/1.22 = 0.378 \text{ lb/(s) (ft}^2)$$

$$\frac{G^2}{\rho_G}\left(\frac{a}{\varepsilon^3}\right)\frac{\psi^2\mu^{0.2}}{\rho_L g_c} = (0.1535)\left(\frac{0.593}{0.925}\right)^2 = 0.063$$

This indicates operation in the loading region. The expected pressure drop is 0.5 in. H_2O/ft.

Total expected pressure drop:

$$\text{Packing} = (0.5)\ (15) = 7.5 \text{ in. } H_2O$$

Support = 1.5 in. (estimated from Figures 20.29, 20.30, 20.31, and 20.32 for a 58% open grid).

Total drop = 9.0 in. H_2O (approximate)

Superficial gas velocity through tower:

$$= \frac{0.593 \text{ lb/(sec)(ft}^2)}{0.087 \text{ lb/ft}^3} = 6.8 \text{ft/s}.$$

Entrainment

This velocity is slightly high and an entrainment knockout or separator should be installed in the air stream following the tower or on the top of the tower itself.

Liquid Hold-up in the Tower:

For water, the hold-up would be from Equation 20.87.

$$h_{tw} = 0.0004\left(\frac{L'}{d_p}\right)^{0.6} \tag{20.87}$$

$$d_p = 0.68 \text{ (from the Table 20.7)}$$

$$h_{tw} = 0.0004\frac{[(0.0378)(3600)]^{0.6}}{(0.68)} = 0.0384\ \text{ft}^3/\text{cu ft for } H_2O$$

For sulfuric acid:

From Figure 20.6C, h_o/h_{ow} for density correction multiplier = 0.6

From Figure 20.38B, correction for surface tension = 1.0 (at 70 dynes/cm)

From Figure 20.38D, correction for viscosity = 1.1 (at 18 cP)

h_o, for acid = h_{ow} (0.6) (1.0) (1.1) = (0.0384) (0.66) = 0.0254 ft^3 acid/ft tower volume

For a packed volume of 15ft in a 15-in. I.D. tower, the total acid hold-up:

$$= [(15)\ (1.22)]\ (0.0254)\ (112.6\ \text{lb/ft}^3)$$

Total hold-up = 52.3 lb acid

Weights

Weight of dry packing in tower:

$$= (42\ \text{lb/ft}^3)\ [(15)\ (1.22)] = 770\ \text{lb}$$

Total weight on bottom support plate when operating (not flooded)

$$= 52.3 + 770 = 822.3\ \text{lb}$$

Some allowance should be made for surging or uneven operation.

The maximum expected weight of liquid would be at flooding conditions:

Using percent free gas space = 77.5

Volume of liquid space = (15) (1.22) (0.775) = 14.2 ft^3

Weight of acid in this space = (14.2) (112.6) = 1,600 lb

Maximum support load = 770 + 1,600 = 2,370 lb

This is the load that should be considered for the support design and selection. To allow for unusual conditions, specify support load = (1.1) (2,370) = 2,600-lb minimum.

20.27 Structured Packing

Structured packings as in use at the present time are composed of

1. Wire-mesh weavings (Figures 20.6Y, 20.6Z, 20.6AA–20.6FF).
2. Corrugated sheet or crimped sheet (usually somewhat thin) (Figures 20.6GG–20.6NN).
3. Grid-type, open, heavy (usually metal) bar-grid shapes stacked together (Figures 20.6OO–20.6TT).

The preferred process application for structured packings depends on the geometric arrangement of the components and

1. High cost/ft^3 of space
2. Easy plugging by solid particles.
3. Importance of uniform liquid distribution across the packing.
4. System operating under pressure [121] or vacuum.

For most efficient performance, the fabricated sections of the structured packing usually are placed in a specific rotated pattern in the column to ensure uniform liquid flow and vapor cross-mixing.

The materials of construction are usually stainless steel or other specific metals that will draw into wire or crimp without cracking. The wire mesh types are also fabricated from some plastics such as Teflon®, polypropylene, etc.; however, the surface must be wettable by the liquid, or the efficiency will be poor. Performance data are needed to complete a good design.

Nutter [51] reports that most, if not all, structured packings (Figures 20.6LL and 20.6MM) follow the linear relationship of vapor rate vs. pressure drop at fixed liquid rates as exhibited by random packings.

Structured packing ACS ST-100 (Figure 20.6DD) is reported by the manufacturer to be equivalent in all respects to the original Sulzer packing, i.e., extremely low-pressure drop and high efficiency (low HETP) over a wide range of liquid load and high-vacuum applications. The design rating procedure used by this manufacturer is confidential (as is also the best and final procedure for most other structured and grid packings), and design application data should be submitted for recommendations.

Preliminary Sizing for ACS Industries Series Woven X/S Knitted Wire Mesh Structured Packing

The following is extracted by permission from ACS Bulletin B-129 (Apr. 1992). Typical HETP is 3 in. compared to ACS products ST-100 of 7 in. Approximate comparison data are ΔP as low as 0.1 in. water/theoretical tray and turndown ratio about 20:1. Special series S construction is recommended for smaller columns, and transverse Series X for larger ones. The suggested preliminary sizing procedure is:

1. Determine the number of theoretical plates/trays/stages using standardized methods.
2. Calculate the minimum allowable superficial vapor velocity (flooding) for equal vapor and liquid loads, for the X-100 and S-100 packings.

$$u_{\max IE} = [0.0942/\mu^{0.33}][(\rho_L - \rho_V)/\rho_V]^{0.57}, \text{ft/s.} \tag{20.88}$$

where

μ = liquid viscosity, cP
ρ_L = liquid density, lb/ft^3.
ρ_V = vapor density, lb/ft^3.

If liquid viscosity is less than 0.15 cP, use constant 0.53 in place of the $\mu^{0.33}$ term.

$u_{\text{max IE}}$ = calculated or estimated maximum superficial vapor velocity.

Calculate: L/G

L = total liquid load, lb/h.

G = total vapor load, lb/h.

Use Figure 20.40 with this L/G and find the load ratio correction factor. Multiply $u_{\text{max IE}}$ by this factor to determine the corrected maximum superficial vapor velocity, $u_{\text{max 1}}$, for X-100 and S-100 packings.

Correct the maximum vapor velocity again for the packing type, if other than X-100 or S-100

$$u_{\text{max}} = u_{\text{max1}}(SS_1\,PV_A/SS_A\,PV_1)^{0.5} \tag{20.89}$$

where

u_{max} = maximum vapor velocity for actual packing, ft/s.

$u_{\text{max 1}}$ = maximum superficial vapor velocity for X-100 and S-100 packing.

SS_A = specific surface for the actual packing* (see Table 20.38A–C).

SS_1 = specific surface for X-100 and S-100 packings (see Table 20.38).

PV_A = percent void volume for the actual packing* (see Table 20.38).

PV_1 = percent void volume for the X-100 and S-100 packings.

Calculate required column area and diameter using the maximum superficial vapor velocity $u_{\text{max 1}}$ and the actual vapor load.

Convert the mass vapor load, G, (lb/h) to the actual vapor volumetric flow, V, in ft³/s. Divide by u_{max} to obtain the minimum cross-sectional column area, A, ft².

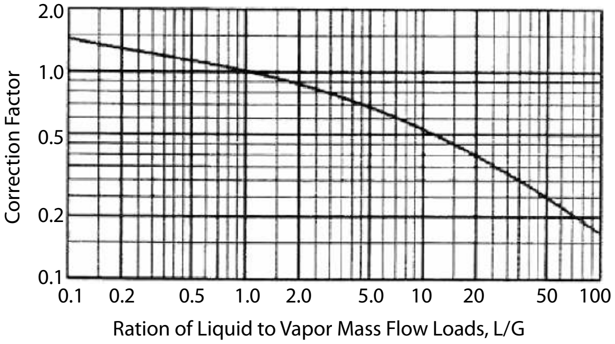


Figure 20.40 ACS maximum vapor velocity correction for L/G for woven/knitted wire mesh structured packing. Used by permission of ACS Industries, Inc., Separation Technology Division, Bull. B-129 (1992).

Table 20.38A JPG mechanical characteristics of common standard packing styles (ACS Wire Mesh).

ACS packing style	Number of strands	Strands diam., in.	Density, lb/ft³ (stainless)	Pct. void vol.	Sp. surf., ft²/ft³
X-100, S-100*	12	0.0045	27.5	94.5	585
X-200*, S-200	8	0.0045	20.1	96.0	426

*For Series X, standard style is X-200 (8 strands)

For Series S, standard style is S-100 (12 strands)

By permission of ACS Industries, Separations Technology Group, Bul. B-129 (1992).

Table 20.38B Experimental HETP values from various sources for Packings Equivalent to Stainless Steel X-200.

System	Diam., in.	Absolute mm Hg	HETP, in.
Methylcyclohexane and toluene	1.25 1.25	760 200	1.25–2.00 1.75–2.50
Methylcyclohexane and toluene	6 18	200–760 200–760	2–3 4–6
Improved liquid distribution →	18	200–760	3.5
Methylcyclohexane and heptane	3	760	1.25–2
Ethylene dichloride and benzene	12	760	4–9
Orthodichlorobenzene and orthodiethylbenzene	1.25 1.25	50 10	1.25 1.65
Phenol and orthocresol (no distributor used)	4 4 4	7–25 85–100 760	2.5–4 2–3.5 2–3.5
n-Hexanol and aniline	1.25	50–300	2.5–2.75
Nitrobenzene and aniline	1.25	5	3.5–4
n-Decanol and methyl naphthalene (no distributor used)	4	3	5
n-Decanol and methyl naphthalene Methanol and water	4 1.25	3 760	4 9

By permission of ACS Industries, Separations Technology Group, Bul. B-129 (1992).

Table 20.38C Rough HETP correction factor for diameter.

Column diameter	HETP proportional to:
Below 2 in.	1.0
2 to 6 in.	1.5
6 to 18 in.	2.0
Over 18 in.	2.5 to 3.0

By permission of ACS Industries, Separations Technology Group, Bul. B-129 (1992).

Correct this area by dividing the fraction of capacity at which the designer intends for the column to operate. A value of 0.7 is recommended, unless some other consideration suggests a different percentage.

$$A = V/u_{max} = \text{column area, ft}^2. \tag{20.90}$$

Final column inside net area = $A_F = A/0.70$ capacity factor

Calculate the column net inside diameter from the area values. Round to the nearest practical commercial column diameter such as 12 in., 15 in., 18 in., 24 in., 30 in., 36 in., and above in increments of 6 in. Then recalculate the actual resulting vapor velocity.

$$\text{or, } A_o = A/0.7, \text{ft}^2 \text{ required} \tag{20.91}$$

$$\text{Column diameter} = \sqrt{4A_o/\pi}, \text{ft}. \tag{20.92}$$

Commercial size columns using Series X or Series S packing will operate at full efficiency with liquid loads as low as 5 gal/h/ft^2 column area, and sometimes less. They can operate satisfactorily exceeding liquid rates of 2,650 gal/h/ft^2. See Figure 20.41.

HETP for ACS Series X-200 Structured Packing:

1. If plant or pilot plant data are available, use them.
2. Table 20.38B presents references for packings equivalent to the X-200 for the systems and conditions noted. See Figure 20.42.
3. Using an available HETP value for a specific Type X or S packing to convert to a column of different diameter, use the "rough" correction factors in Table 20.38C. These factors recognize that larger diameter columns have more uneven liquid distribution. When no HETP values or data are available, it is suggested to use the approximation developed for S-100 packing.

Estimating HETP.

$$\mathrm{HETP} = [0.18 + \mu^{0.55}]/[1.23 + (3.15/R^{0.27})] \tag{20.93}$$

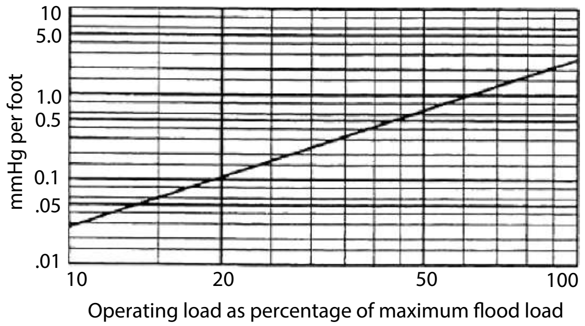


Figure 20.41 Estimated pressure drop referenced to percentage of maximum flood load for structured woven/knitted wire mesh X-100 and S-100. Referenced to liquid of 60 lb/ft^3. To correct other liquids, multiply ΔP from figure by ratio of actual liquid density to 60. Used by permission of ACS Industries, Inc., Separation Technology Division of ACS Industries, Inc. Separation Technology Division, Bull. B-129 (1992).

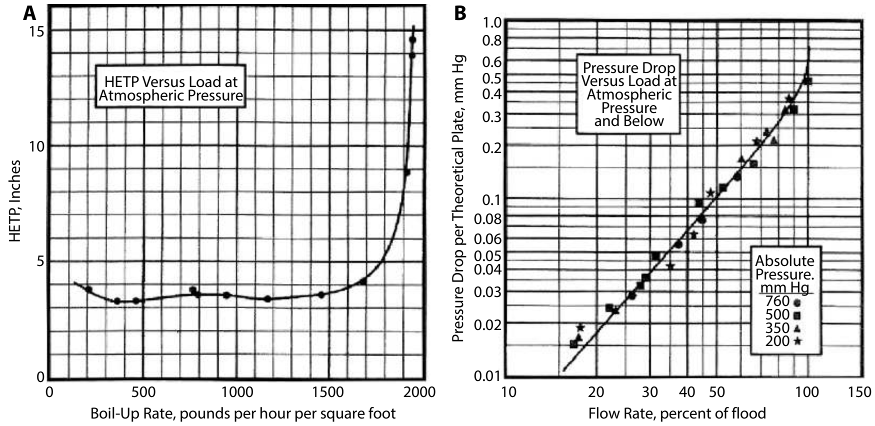


Figure 20.42 HETP and pressure drop data for a typical distillation system. Packing equivalent to X- 200 (eight strands), stainless steel. System: methylcyclohexane and toluene. Reflux ratio: 100% Column Diameter: 18 in. Packed height: 5 ft. Used by permission: ACS Industries, Inc., Separation Technology Division, Bull. B-129 (1992).

where

HETP = height equivalent to a theoretical plate/tray/stage, ft.
μ = liquid viscosity, cP
R = liquid rate, gal/h/ft^2

If $\mu < 0.30$ cP, use 0.52 in place of the $\mu^{0.55}$ term. To convert an available HETP value for a given system and column diameter to a different packing in the same series (X or S), assume HETP is inversely proportional to the specific surface (consult manufacturer for this detail).

$$\mathrm{HETP_2 = HETP_1(SS_1/SS_2)} \tag{20.94}$$

where 1 and 2 = two different wire mesh packings.

For highly varying vapor loads through the column, pressure drop may be estimated for differences in latent heats from one point in the column to another; the pressure drop should be calculated for each foot of packing, or less adjusting the load. This particularly applies to vacuum columns or tall columns at high pressures. Maximum heights of a packed section in a column should be determined after consultation with the manufacturer.

Pressure Drop (Estimated)

Use Figure 20.41 to estimate column pressure per foot of packing height, referenced to liquid load as percentage of maximum flood load. For varying load, perhaps ±20%, vacuum systems and varying latent heats in the column, calculate the pressure drop for each foot of packing or less; this requires adjusting the percentage load in each calculation.

Koch Kulzer Structured Packing [66]

This packing is woven wire mesh with the following characteristics as reported by the manufacturer/licensee [66]:

1. HETP of about 7 in. even at low liquid and vapor loads.
2. Efficiency essentially independent of column diameter.
3. Capacity and pressure drop characteristics superior to the best commercially available mass transfer devices.
4. Specific pressure drop of 0.2 mm Hg per theoretical tray.

This packing is a woven wire fabric of parallel corrugated elements and is completely self-wetting. It does not require pre-flooding to attain the wetted condition. This packing has a flat surface of approximately 150 ft^2/ft^3 and a free area > 90%.

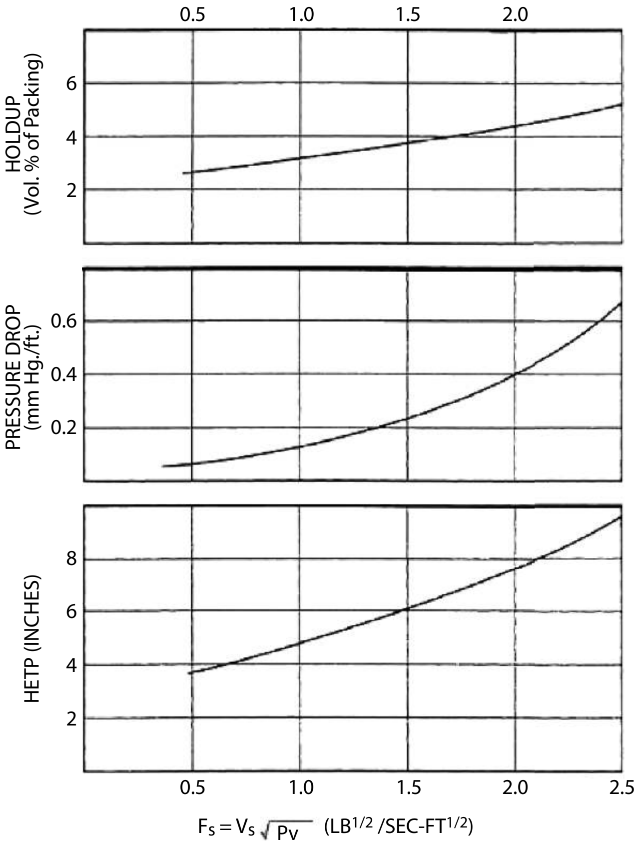
In operation, the liquid flows downwards in a zigzag pattern. See Figure 20.6EE.

Example 20.8: Koch-Sulzer Packing Tower Sizing (used by permission, Bulletin KS-1, Koch Engineering Co. Inc.)

The following typical design problem illustrates the calculation procedure for Koch Sulzer Packing. A comparison of a KS Packing design and a Flexi ring design will result in (for 2-in metal, Pall rings) top = 7.5-ft. D and bottom = 8.5-ft D. See Figures 20.43 to 20.46.

Distill a diethanolamine (DEA)–triethanolamine (TEA) mixture to produce a 99.0 wt % DEA distillate product and a 95.0 wt % TEA bottoms product. The design material balance is as follows:

	Feed		**Distillate**		**Bottoms**	
	lb/mol	**mol/h**	**lb/mol**	**mol/h**	**lb/h.**	**mol/h**
DEA	5,000	47.6	4,740	45.1	260	2.5
TEA	5,000	33.5	50	0.3	4,950	33.2
Totals	10,000	81.1	4,790	45.4	5,210	35.7



Column Diameter .... 20 inches
System .... cis-trans-decalin
Top Pressure .... 20 mm Hg.
Reflux Ratio .... Infinite

Figure 20.43 Characteristics of Koch/Sulzer packing. Gas loading factor, F versus HETP, pressure drop, and packing hold-up. Note: Vs = superficial gas velocity, ft/s and ρ_v = vapor density, lb/ft^3. Used by permission of Koch Engineering Co., Inc., Bull. KS-1 and KS-2.

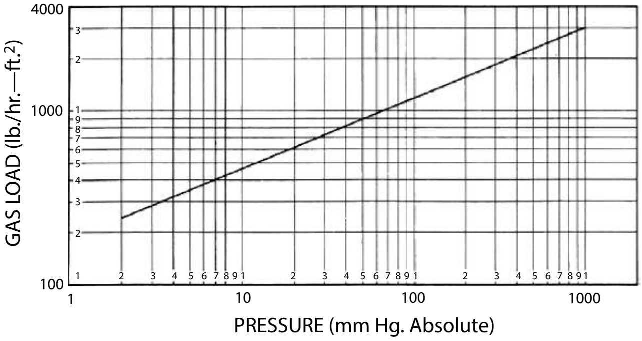


Figure 20.44 Koch/Sulzer Design mass velocity (lb/h-ft^2) for HETP of 7 in. System cis-trans-decalin, molecular weight = 138. Used by permission of Koch Engineering Co., Inc., Bull. KS-1.

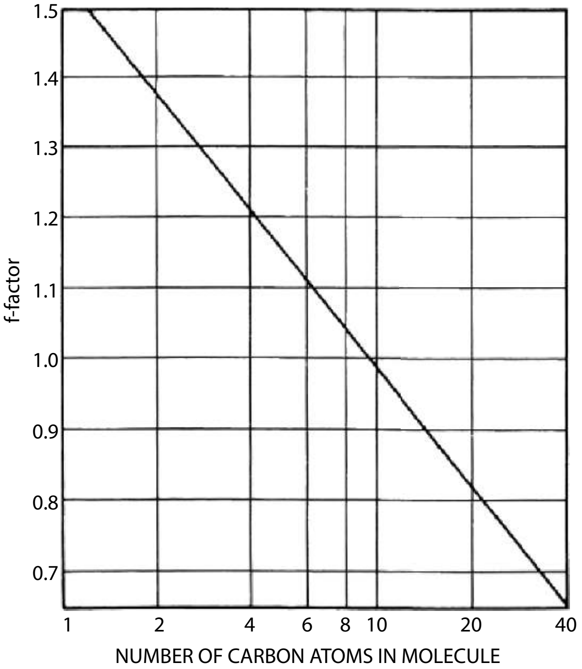


Figure 20.45 Koch/Sulzer packing. Design mass velocities for systems other than cis-trans-decalin are obtained by dividing the values from Figure 20.44 by the "f" factor. For the following groups corrections should be applied to the number of carbons in the molecule:

Paraffins, olefins acids, aldehydes, ketones	0
Benzene rings, alcohol, phenol	−1.0
Saturated ring	−0.5
Nitro	+1.0
Ether	+1.5
Ester, secondary amine	+2.0
Chlorine	+3.0
Bromine	+8.0

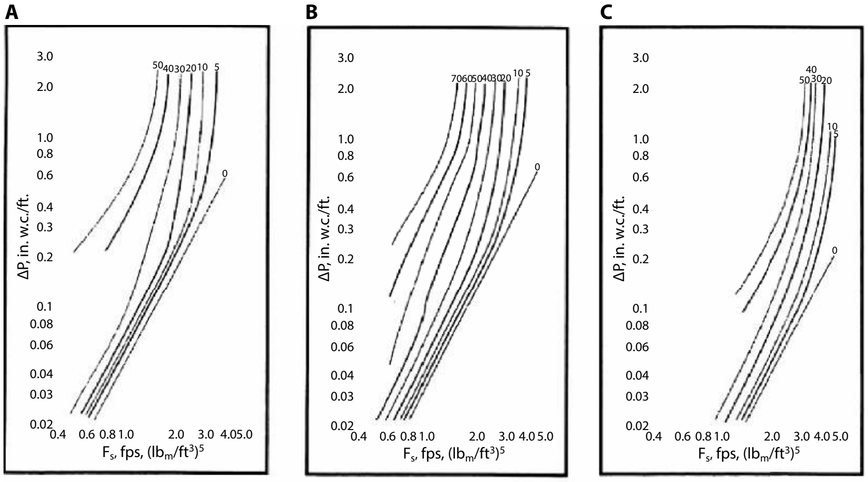
Used by permission of Koch Engineering Co., Inc., Bull. KS-1.

To preclude thermal degradation of the amine solution, the reboiler pressure must be limited to 10 mm Hg absolute pressure. Total allowable pressure drop is 3 mm Hg.

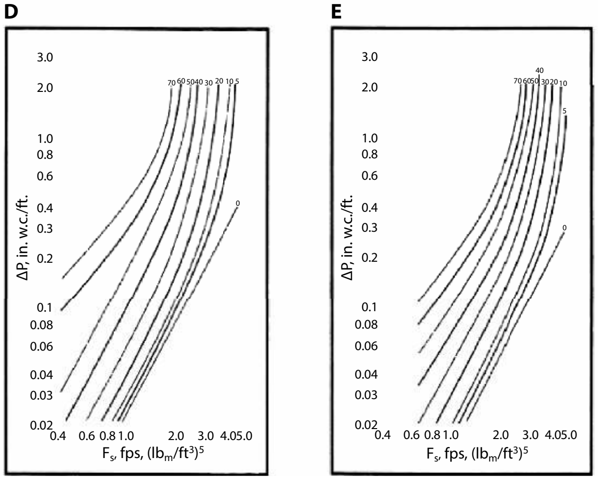
Solution

Distillation calculations result in a reflux ratio L/D = 0.8, with 4 theoretical stages/trays for rectification and 4 theoretical trays for stripping, or a total of 8 trays. The design heat balance (neglecting heat losses) is as follows:

Heat Out	lb/h	State	°F	Btu/lb	Btu/h.
Distillate	4,790	Liquid	296	160	766,000
Bottoms	5,210	Liquid	381	240	1,250,000
Ovhd. Cond.	–	–	–	–	2,470,000
	10,000				4,486,000



Pressure drop for Type 2Y FLEXIPAC® Pressure drop for Type 2.5Y FLEXIPAC® Pressure drop for Type 3X FLEXIPAC®



Pressure drop for Type 3Y FLEXIPAC® Pressure drop for Type 4Y FLEXIPAC®

Figure 20.46 Typical pressure drop-capacity curves for counter current gas–liquid operation, for Koch Flexipac®. Used by permission of Koch Engineering Co., Inc., Bull. KFP-4. Parameter lines on charts are gpm/ft^2.

Heat In					
Feed	10,000	Liquid	314	180	1,800,000
Reboiler (Q_r)		By Difference			2,686,000
	10,000				4,486,000

Column Sizing for Koch Sulzer Packing

Bottom Section

The bottom section vapor loading is controlled by the reboiler vapor where

$$V_r = Q_r / l_v = TEA = 2{,}686{,}000/192 = 14{,}000 \text{lb/h}$$

The design mass velocity at 10 mm Hg (Figure 20.44) is 460 lb/h-ft^2. Applying the "f" factor correction for six carbon atoms (Figure 20.45) results in

$$f = 1.12$$

$$A = 14{,}000\ (1.12)/460 = 34.1\ \text{ft}^2$$

$$D = 6.6\ \text{ft, use 6 ft. 9 in.}$$

$$\begin{aligned} V_s &= V_r/3600(35.8)\rho_V \\ &= 14{,}000/3{,}600(35.8)(0.0031) = 35\ \text{ft/s.} \end{aligned}$$

$$F_s = V_s(\rho_V)^{1/2} = 35(0.0557) = 1.95\ \text{lb}^{1/2}/\text{ft}^{1/2}\ \text{s}$$

Considering the reboiler as a theoretical tray, for three theoretical trays, the packing height for an HETP of 7 in. is 7 (3)/12 = 1.75 ft. This requires four standard elements, which is 6.7 in. deep, or 80 ft^3 of packing.

From Figure 20.43, the pressure drop is 0.38 mm Hg/ft or 0.85 mm Hg. total.

Top Section

The top vapor load at an L/D of 0.8 is 8,620 lb/h. The total pressure drop for seven theoretical trays is slightly less than 2 mm Hg and ρ_V is 0.002 lb/ft^3 at 8 mm Hg top pressure. Duplicating the calculations made for the bottom section results in

D = 5 ft. 9 in.
Packing height = 2.8 ft.
Pressure drop = 1.06 mm Hg.
Therefore, use column diameter = 6 ft. 9 in.

Nomenclature

A = cross-sectional area, ft^2
D = diameter, ft.
F_s = $V_s(\rho_V)^{1/2}$
G = vapor rate, lb/sec. ft^2
g_c = 32.174 (lb_m/lb_f) (ft/s^2)
L = Liquid rate, lb/s. ft^2
L/D = reflux ratio.
P_f = packing factor
Q_c = condenser duty, Btu/h.
Q_r = reboiler duty, Btu/h.
V_r = reboiler vapor rate, lb/h.
V_s = superficial vapor velocity, ft/s.
ρ_1 = liquid density, lb/ft^3
ρ_V = vapor density, lb/ft^3
l_v = latent heat of vaporization, Btu/lb

Koch Flexipac® Structured Packing

This comes in four sizes, Types 1 through 4, and is constructed of corrugated metal sheets (see Figure 20.6GG). The types vary by corrugation size; the larger the type number, the greater is the depth of corrugation. The deeper corrugations give higher capacity and lower pressure drop. According to Koch reference [67], at the same efficiency, in countercurrent gas–liquid operation, this packing has a higher capacity and lower pressure drop than any available dumped or structured packing. The terminology for Figure 20.46 is

$$F_s = V\sqrt{\rho_V} = [G/3{,}600(\rho_V)(A)]\left(\sqrt{\rho_V}\right), \text{ft/sec}. \tag{20.95}$$

where

A = cross-sectional area, ft^2
G = vapor rate, lb/h.
V = vapor velocity, ft/s.
ρ_V = vapor density, lb/ft^3
ΔP = pressure drop, in. H_2O/ft height.

Chart parameter lines are gpm/ft^2 cross-section.

The Flexipac® structural packings have better efficiency than available random packing, particularly at low liquid rate; see Reference [67].

Intalox High-Performance Metal Structural Packing [68]

According to the manufacturer's literature [68], this packing surpasses the best of other sheet-metal structured packings in terms of efficiency and capacity. See Figure 20.6II. The unique surface-texturing feature provides greater use of the packing surface to achieve enhanced levels of mass transfer, and the overall geometry allows greater capacities and efficiencies to be obtained. Tests have been conducted on this and other packings at the University of Texas at Austin's "Separation Research Program, Center for Energy Studies" for distillation capacity and efficiency, and published in Reference 69.

For good and uniform performance of any structured packing, it is essential to have uniform, consistent vapor and liquid distribution; therefore, much care must be given to the design details. See earlier discussion in this chapter.

For specific final performance sizing of a distillation column using Norton's Intalox® structured packing, the designer is referred to the manufacturer's technical representatives, and should not assume that the preliminary results obtained from any manufacturer's bulletin that are included here will necessarily serve as a final design. As a preliminary examination of a design problem (used by permission of Norton Chemical Process Products):

1. Calculate flow parameter X

$$X = (L/G)\sqrt{\rho_G/\rho_L} \tag{20.96}$$

where

A = area, ft^2 tower cross-sectional area
L = Liquid mass rate, lb/s.
G = vapor or gas mass rate, lb/s.
ρ_G = gas density, lb/ft^3 at conditions
ρ_L = liquid density, lb/ft^3 at conditions

2. Read chart, Figure 20.47, and obtain C_o, ft/s, at packing type shown.

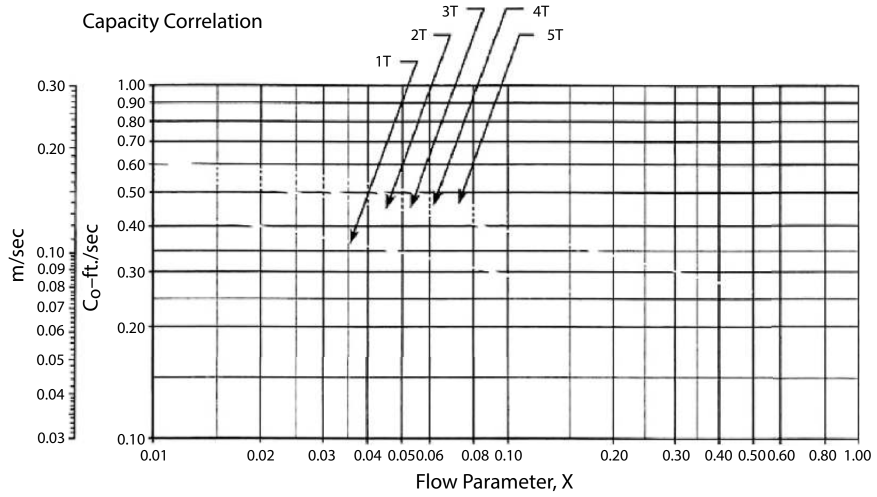


Figure 20.47 Capacity correlation for two types of Intalox® structured packing. Data range $5 \le \sigma \le 73$ and $0.07 \le \mu \le 1.1$. Used by permission of Norton Chemical Process Products Corp. Bull. ISP-2 (1994).

3. Calculate efficient capacity, C_{SC}:

$$C_{SC} = C_o[\sigma/20]^{0.16}[\mu/0.2]^{-0.11} \tag{20.97}$$

σ = surface tension, dynes/cm.
μ = liquid viscosity, cP

4. Calculate

$$C_S = V\sqrt{\rho_G/(\rho_L - \rho_G)},\ \text{ft/s.}$$

V = superficial gas velocity, ft/s., or m/s. depending on the units used

$$V = G/(\rho_G A),\ \text{ft/s.}$$

5. Read pressure drop coefficient, F
   Read, F, for the approximate calculated C_S, ft/s, from Table 20.39 at calculated liquid rate, lb/h.ft². Then read Y from IMTP Packing Pressure Drop chart, Figure 20.13G, and then read curves showing pressure drop (may require interpolation).
6. Typical HETP data are shown in Figures 20.48A to 20.48C.

Table 20.39 Norton Intalox® structured packing pressure drop coefficient F* for type 2T packing.

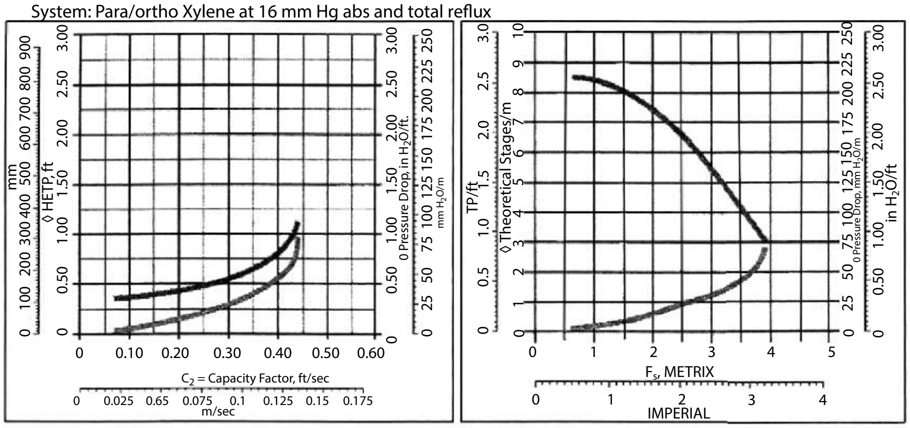
Air/Water	Liquid, kg/h · m²	9,760	24,400	48,800	73,200	97,600	122,000
	Liquid, lb/h · ft²	2,000	5,000	10,000	15,000	20,000	25,000
	Intalox structured packing 2T						
	F when C_s in m/s	205	215	250	300	325	365
	F when C_s in ft/s	19	20	23	28	30	34

These packing factors can be used to predict the pressure drop of Intalox Structured Packing 2T when used with the Generalized Pressure Drop.
Correlation as illustrated in Norton's Intalox High-Performance Separation Systems; Figure 14.21G this text.
Used by permission of Norton Chemical Process Products Corp., Bull. I-S-I-R.

A

Intalox® Wire Guaze Packing

System: Para/ortho Xylene at 16 mm Hg abs and total reflux



Intalox Structured Packing 1T

System: Iso-octane/Toluene at 100 mm Hg abs and total reflux

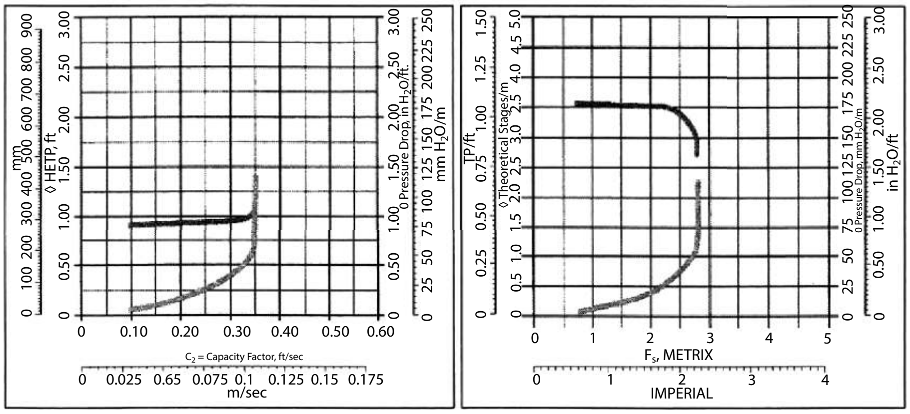


Figure 20.48A Performance test results using Norton's Intalox® structured packing. Used by permission of Norton Chemical Process Products Corp., Bull. ISP-2.

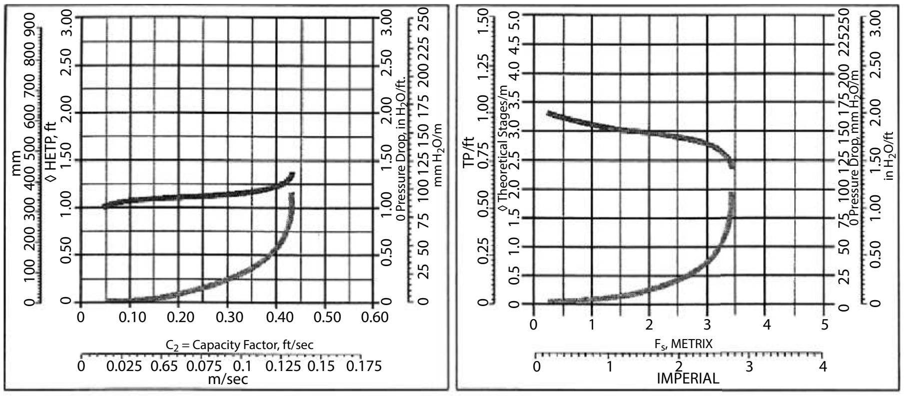
Gempack® Structured Packing; Glitsch, Inc.

These packings are represented in part by Glitsch, Inc's bulletin [70] (Figure 20.6KK) and they cite the performance of the packing as

1. High efficiency: 5–25 in. HETP (7.9–1.6 theoretical plates per meter).
2. Low-pressure drop: 0.5–0.7 mm Hg/theoretical plate.
3. High vapor loadings: up to 0.5 C-Factor (up to 4.0 F-factor).
4. High liquid capacity: 0.5–55 U.S gpm/ft^2 (1.22–134 m^3/m^2/h).
5. Wide turndown ratio: limited only by distributors
6. Availability: diameters from 4 in. to over 45 ft (10 cm to 13.7 meters).
7. Series available:
   - (a) Series AT: General purpose, suitable for low wetting rates and high vacuum applications.
   - (b) Series AS: General purpose, suitable for high wetting and heat transfer applications.
   - (c) Series AW: Latest generation of Gempack®, better efficiency with same hydraulic capacity as other series listed.
   - (d) Series AL: Especially suited for low wetting rates and chemical applications.
   - (e) Series BG: Wire gauze packing, crimped to 60° from horizontal, for high efficiency applications and very low wetting rates in clean service.

B **Intalox Structured Packing 2T**

System: Iso-octane/Toluene at 100 mm Hg abs and total reflux



Intalox Structured Packing 3T

System: Iso-octane/Toluene at 100 mm Hg abs and total reflux

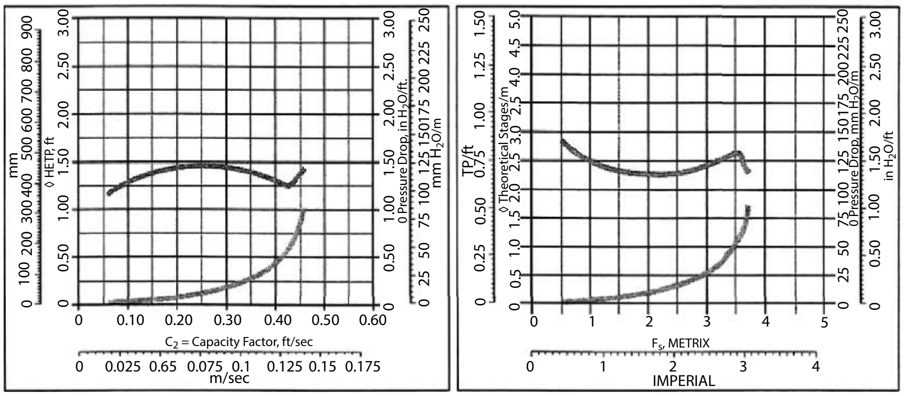


Figure 20.48B Performance test results using Norton's Intalox® structured packing. Used by permission of Norton Chemical Process Products Corp., Bull. ISP-2.

The performance and design information in Glitsch reference [70], as for all the other manufacturers with respect to either data or charts, is proprietary, but not necessarily warranted to be suitable for the designer's service/applications unless verified by the manufacturer's representatives.

Example 20.9: Heavy Gas-Oil Fractionation of a Crude Tower Using Glitsch's Gempak® (Used by permission of Glitsch, Inc. Bulletin 5140)

The heavy naphtha-light gas-oil fractionation zone of a crude tower has to be revamped to handle 25% more capacity. Because trays would be working at high percent flooding. Gempak structured packing is considered (Figures 20.48A–C).

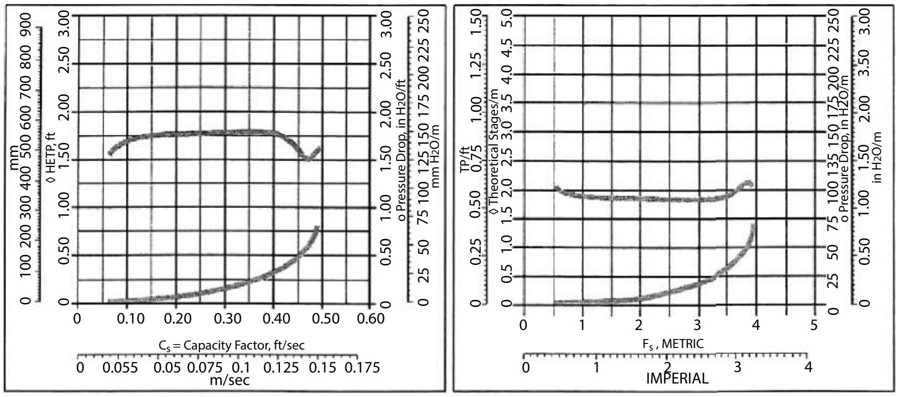
The loads are as follows:

Vapor rate	688, 000 lb/h
Liquid rate	381,000 lb/h
Vapor density, (ρ_V)	0.355 lb/ft^3
Specific gravity of gas oil	0.68
Liquid density, (ρ_L)	42.42 lb/ft^3
Liquid viscosity,	0.281 cP
Non-Foaming System	

C

Intalox Structured Packing 4T

System: Iso-octane/Toluene at 100 mm Hg abs and total reflux



Intalox Structured Packing 5T

System: Iso-octane/Toluene at 100 mm Hg abs and total reflux

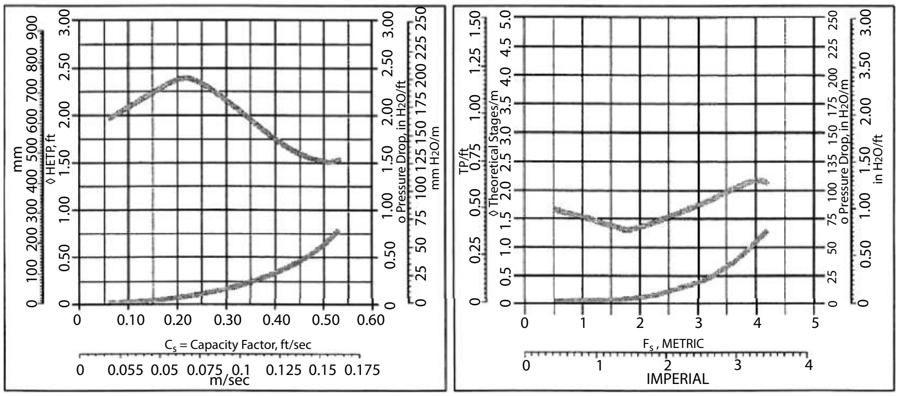


Figure 20.48C Performance test results using Norton's Intalox® structured packing. Used by permission of Norton Chemical Process Products Corp., Bull. ISP-2.

Tower diameter:	15 ft – 0 in.
Cross-sectional area	176.72 ft^2
Height available for packed bed (excluding height for distributor)	6.5 ft (78 in.)

Solution

1. Calculate the vapor load:

$$\text{Vapor, } (688{,}000)/(3{,}600)\,(0.355) = 538.3 \text{ ft}^3/\text{s.}$$

$$\text{Vapor velocity} = V_s = (538.2)/(176.72) = 3.046 \text{ ft/s.}$$

$$\begin{aligned}\text{C-factor} &= V_s[(\rho_V)/(\rho_L - \rho_V)]^{1/2} \\ &= 3.046[(0.355)/(42.42 - 0.355)]^{1/2} = 0.28\end{aligned}$$

2. Calculate the liquid load:

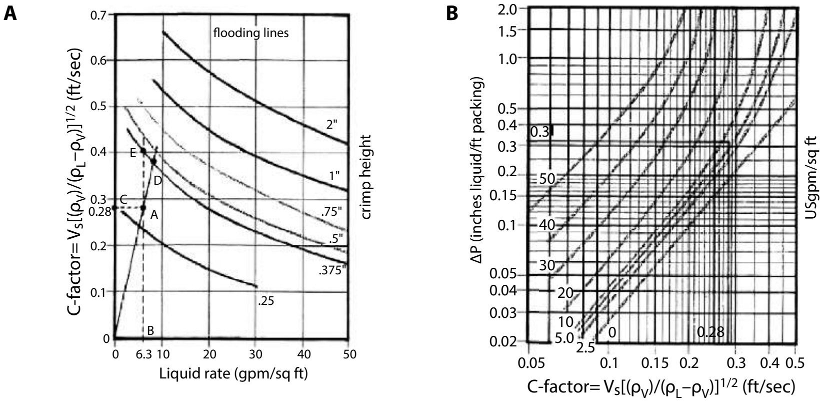
$$\text{gpm} = (381{,}000)/(8.33)\,(60)\,(0.68) = 1{,}120$$

$$\text{gpm/ft}^2 = (1{,}120)/(176.72) = 6.3$$

3. Select the Gempak size:
   On the plot of C-factor vs. liquid load as shown in Figure 20.49A, mark the operating point (A) using 6.3 gpm/ft^2 and C-factor = 0.28. The point on the figure is above the Gempak AS (0.25-in. crimp height) packing flooding line. All other packing sizes would be operating below flooding. For example, if Gempak AS (0.375-in. crimp height) is to be used, the percent flooding at design loads will be

   - At constant L/V ratio – (OA/OD) x 100 = 75.2%
   - At constant liquid rate – (100) (BA/BE) = 70.0%

   Normally in distillation, flooding at constant L/V ratio is more representative of actual plant operations (constant liquid loads may also be representative in cases like absorbers). In general, percent flooding up to 75–80% is acceptable for continuous operation. Gempak AS (0.375-in. crimp height) would be a good selection for this example. Gempak AS (0.5-in. crimp height) would also be a good selection, with flooding of 70% at constant L/V ratio and 65% at constant liquid rate. Gempak AS (0.25-in. crimp height) would not be a good selection in this case since flooding would be over 100%.
4. Pressure Drop
   On pressure drop plots, mark a C-factor of 0.28 and 6.3 gpm/ft^2 as shown in Figures 20.49N and 20.49C.



Gempak® flooding line for Series AS. Used by permission of Glitsch, Inc., Bull. 5140.

Gempak® pressure drop, Series AS, with 0.50-in. crimp height. Used by permission of Glitsch, Inc., Bull. 5140.

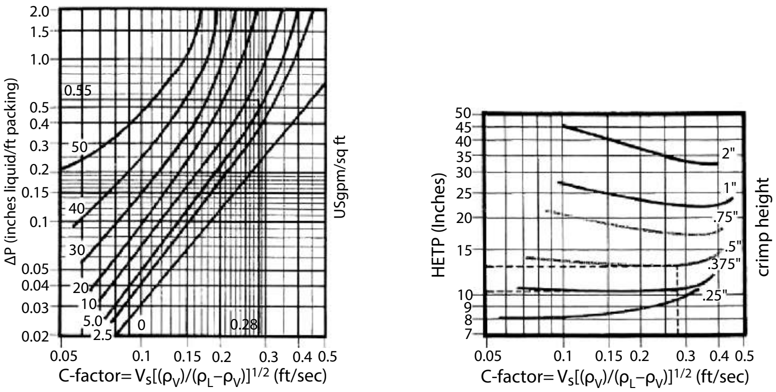


Figure 20.49 For Example 20.8 typical Glitsch Gempak® preliminary design charts for structured packing. Note: These plots are to be used in carrying out the calculations described in the Glitsch bulletin. Graphic data presented here have been obtained in tests whose conditions may differ materially from your own. Curves on these pages are from Gempak with crimp angles of 45° from the horizontal, and should be sufficient for a preliminary review. The actual surface texture, which determines the final design should be chosen in consultation with the Glitsch process engineering staff. Used by permission of Glitsch, Inc. Bul. 5140.

For Gempak AS (0.5-in. crimp height), $\Delta P = 0.31$ in. liq/ft.
For Gempak AS (0.375-in. crimp height), $\Delta P = 0.55$ in. liq/ft.
For a bed of 6.5 ft, the pressure drop for the bed is
For Gempak AS (0.5-in. crimp height) = 0.31 x 6.5 = 2.0 inches of liquid.
For Gempak AS (0.375-in. crimp height) = 0.55 x 6.5 = 3.6 inches of liquid.
Pressure drop for both AS, 0.5 in. crimp height, and AS, 0.375-in. crimp height, packings is acceptable for crude tower operation.

5. Efficiency, Figure 20.49D
   On efficiency plot, mark C-factor at 0.28.
   For Gempak AS (0.5-in. crimp height): HETP = 13.5 in. (0.89 NTS/ft)
   For Gempak AS (0.375-in. crimp height): HETP = 10.2 in. (1.18 NTS/ft)

Note: The HETPs are noted valid only for o-/p xylene at the test conditions. Nevertheless, the ratio of HETPs should remain approximately constant.

Grid Packing: Nutter Engineering (Figure 20.6PP)

For applications and design details, refer to the manufacturer concerning these types of packings. Figure 20.50 illustrates performance of No. 3 Snap Grid™. For mass transfer for distillation HETP, use

$$\mathrm{HETP} = H_{og}(\ln\lambda)/(\lambda - 1) \tag{20.98}$$

where

$$\lambda = m(G_m/L_m) \tag{20.99}$$

HETP = height equivalent to a theoretical plate, inches
H_{og} = height of an overall gas phase transfer unit, inches
U_a = volumetric overall height transfer coefficient, Btu/(h) (ft^3)(°F)
m = slope of equilibrium line expressed in mole fraction.
G_m, L_m = gas, liquid molar rate based on superficial tower area, mol/(h) (ft^2).
H_g, H_l = height of gas, liquid phase transfer unit, inches
Re = Reynolds number of gas phase

Koch Flexigrid® Packing: Koch Engineering Co.

Koch Flexigrid® packing bulletin [71] states that the packing (Figures 20.6SS, 20.6TT) has a fixed, ordered orientation and is supplied as layers that stack in a prescribed fashion within the bed. Features include [71]

1. High capacity; constructed in 60 in. x 16 in. x 2¾ in. high modules.
2. High efficiency: constructed in 60 in. x 16 in. x 2¾ in. high modules.
3. Each successive layer of the grid is rotated 45° during installation.
4. Lower pressure drop.
5. Tendency to coke or foul far less than other grids due to elimination of horizontal planes where liquids or solids can stagnate.
6. Low liquid hold-up.
7. Fabricated of most metals, such as carbon steel, stainless steel, aluminum, and others as required.

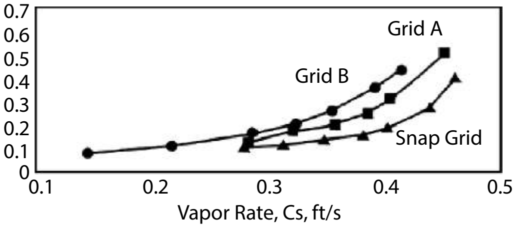


Figure 20.50 Nutter Snap-Grid™ typical performance charts for pressure drop. Used by permission of Nutter Engineering, Harsco Corp., Bull. CSG-2, for Air-Isopar© 10 gpm/ft².

Flexigrid® Style 2 High Capacity

$$\%\ \text{flood}\,@\,\text{const L/V} = 119\times\left(C_v + 0.074\sqrt{C_v\,C_L} + 0.00136C_L\right) \tag{20.100}$$

$$\%\text{flood}\,@\,\text{const L} = \frac{100C_v}{0.84 - 0.0676\sqrt{C_L} + 0.00136C_L} \tag{20.101}$$

Flexigrid® Style 3 High Efficiency

$$\%\ \text{flood}\,@\,\text{const L/V} = 165.34\times\left(C_v + 0.06\sqrt{C_v\,C_L} + 0.009C_L\right) \tag{20.102}$$

$$\%\text{flood}\,@\,\text{const L} = \frac{100C_v}{0.605 - 0.0464\sqrt{C_L} + 0.0009C_L} \tag{20.103}$$

const = constant

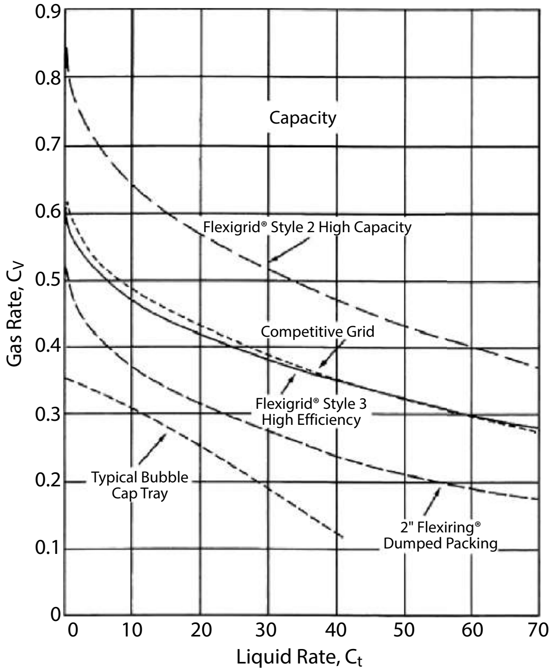


Figure 20.51 Comparison of capacities of Flexigrid® Styles 2 and 3 at flooding with 2-in. Flexring® random packing, and a competitive grid. Used by permission of Koch Engineering Co., Inc., Bull. KFG-2.

A "viscosity correction" should be made if $\mu_L > 10.0$ cP by multiplying the "% flood" obtained from Equations 20.93 through 20.96 by the term "$\mu_L^{0.06}$" in cP.

$$C_v = \frac{V_v}{A}\left(\frac{\rho_V}{\rho_L - \rho_V}\right)^{1/2}, \text{ft/sec}. \tag{20.104}$$

$$C_L = \frac{V_L}{A}\left(\frac{\rho_L}{\rho_L - \rho_V}\right), \text{gpm/ft}^2 \tag{20.105}$$

where

A = tower area, $(\pi D^2/4)$ ft^2
V_V = vapor rate, ft^3/s.
V_L = liquid rate, U.S. gpm
ρ_L = liquid density, lb/ft^3
ρ_V = vapor density, lb/ft^3

Typical pressure drops are shown in Figure 20.52. Mass transfer and heat transfer evaluations should be referred to the manufacturer.

Glitsch-Grid™ [122] (Figure 20.6UU)

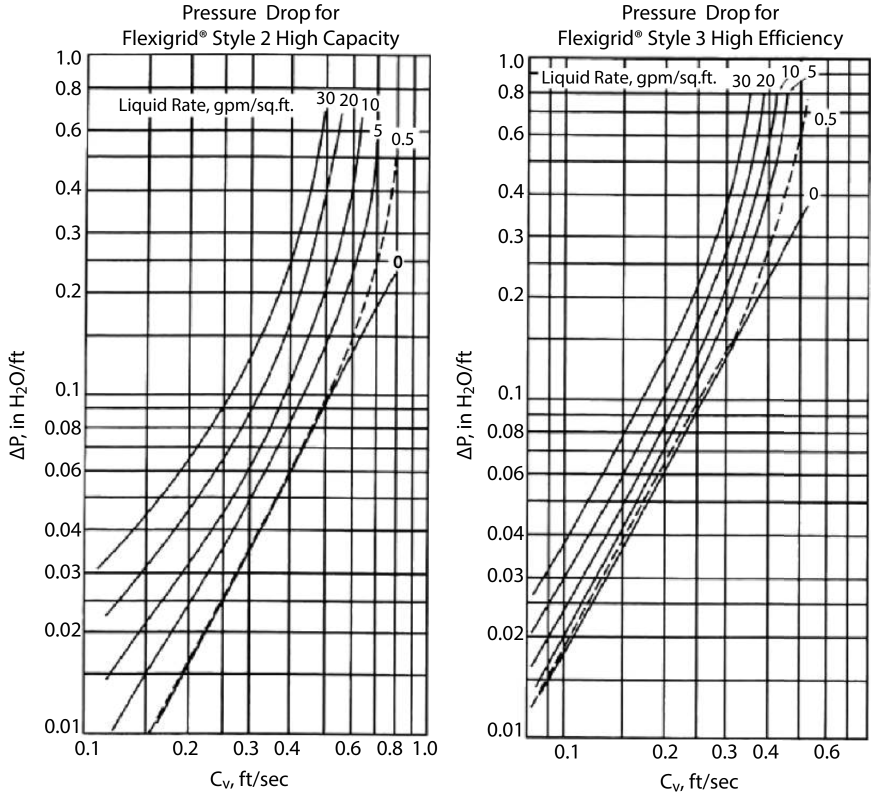


Figure 20.52 Pressure drop for Styles 2 and 3 Flexigrid® at selected liquid rates. Used by permission of Koch Engineering Co., Inc., Bull. KFG-2.

This is an open area packing with multiple layers of lattice-type panels. This grid, as described by the manufacturer's bulletin, consists of vertical, slanted, and horizontal planes of metal. The vertical strips have horizontal flanges oriented alternately right and left. Due to the random overlap, the vapor path must zigzag through the bed.

The manufacturer's data state that the grid has extremely low pressure drop (0.5 mm Hg/ft) at capacities higher than is possible with any other mass transfer device. Grid capacity is approximately 50% greater than conventional trays, and about 35% greater than 3½-in. ballast rings. The grid is highly resistant to fouling, plugging, or coking by tars or solids. See Figures 20.53A and 20.53B for pressure drop and capacity performance comparison. HETP is available from the manufacturer and final design performance must be obtained from the same source.

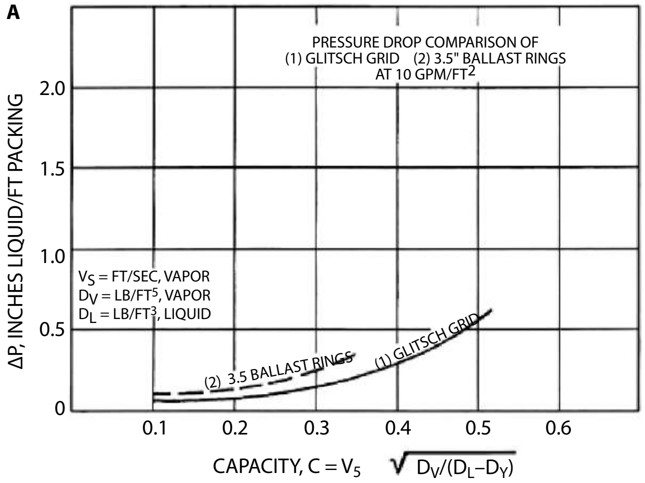


Figure 20.53A Pressure drop comparison of Glitsch Grid™ and Ballast® rings at 10 gpm/ft^2. Used by permission of Glitsch, Inc. Bull. 14-72.

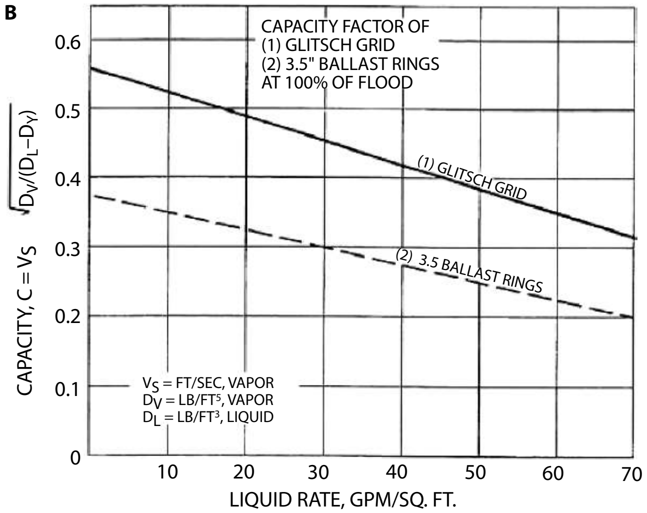


Figure 20.53B Capacity performance comparison for Glitsch – Grid™ and Ballast® rings. Used by permission of Glitsch, Inc., Bull. 14-72.

20.28 Structured Packing: Technical Performance Features

Fair and Bravo [72] have performed extensive studies on structured packing and have developed general models for flooding, pressure drop, and mass transfer. Structured packing is now generally considered cost effective for moderate pressure and vacuum distillation when compared to trays and random packings [72]. The test work of the authors considered the trade-named structured packings of Intalox®, Gempac®, Flexipac®, Mellapak®, Sulzer, and Montz in their studies. See earlier figures for installations of these packings, many of which are quite similar. All of the cited packings are corrugated sheet type designs, except for the Sulzer, which is a fabricated wire gauze construction. Table 20.40 summarizes the characteristics of the selected packings. Refer to the respective manufacturers for confirming details and design application techniques.

Flooding

At flooding or near flooding conditions [72]:

1. A rapidly increasing pressure drop with a relatively slight increase in gas rate (hydraulic flood) develops.
2. A rapidly decreasing efficiency with a relatively slight increase in gas rate (mass-transfer limitation) develops.
3. A general lack of column stability develops.

All of these conditions do not necessarily occur at the same liquid–gas loading. Generally, the mass-transfer limitation develops before the hydraulic flood condition as loadings are increased. Fair and Bravo [72] used the mass-transfer limitation as the limiting case for reasonable design of mass-transfer efficiencies. Figure 20.54 is based on hydraulic flood for several structured packings. The capacity limit is related to the corrugated elements as reflected in specific surface area. The capacity parameter, C_s in m/s, = U_o.

$$C_s = V\sqrt{\rho_G/(\rho_L - \rho_G)},\ \text{m/s.} \tag{20.106}$$

V = superficial velocity, m/s.
ρ_G = gas density, kg/m^3
ρ_L = liquid density, kg/m^3

Table 20.40 Characteristics of representative structured tower packings*.

	Flexipac-2	Gempak 2A	Intalox 2T	Montz B1-200	Mellapak 250Y	**Sulzer BX
Specific area (m^{-1})	223	223	220	200	250	500
Void fraction	0.93	0.95	0.97	0.94	0.95	0.90
Corrugation angle (degrees)	45	45	45	45	45	60
Crimp height (m)	0.0125	0.0122	0.0104	0.0149	0.0119	0.0064
Corrugation side (m)	0.0177	0.0180	0.0223	0.0250	0.0171	0.0088
Corrugation base (m)	0.0250	0.0268	0.0390	0.0399	0.0241	0.0128

*All packing types listed are available in several different sizes. The corrugation angle is measured from the horizontal. Confirm details with manufacturer.
**Wire gauze for comparison.
Used by permission of The American Institute of Chemical Engineers; Fair, J. R. and Bravo, J. L., *Chem. Eng. Prog*. Vol. 89, No. 1 (1990), p. 19; all rights reserved.

Figures 20.55A and 20.55B illustrate for a specific packing the hydraulic flood and mass-transfer efficiency limitations. The differences in crimp height can influence the results. Figure 20.55B shows the effect of a higher flow parameter taken using larger columns; the system apparently was approaching critical, but the cause of the performance is not yet known.

Pressure Drop

Structured packings maintain mass-transfer performance with minimum pressure drop [72]. Two models are presented for calculating pressure drop: (1) Bravo–Rocha–Fair [73] and (2) Stichlmair–Bravo–Fair [74]. Each method is quite involved with rather complex equations to calculate ultimate pressure drop. The authors [72] recommend using the following in design:

1. Flooding: Empirical plots of data, as shown in Figures 20.54, 20.55A, 20.55B, and 20.56.
2. Pressure drop: Generalized model of Bravo *et al.* [73], in combination with Stichlmair [74]. Pressure drop at the flood point as a function of the flow parameter is given in Figure 20.54 [72] for several packings using the Stichlmair model equations for various system pressures [72].
3. Mass transfer: Generalized model of Bravo *et al.* [75], with discount factors for wetted surfaces.

Hatfield [76] describes the improvements in commercial performance when Pall rings were replaced with Goodloe® packing.

Pressure drop through gauze and sheet metal structured packings [77] applies for the region below the loading point and cannot predict the flood point because liquid holdup vs. gas velocity is not included. The latest version of the equation is in Reference 72:

$$\Delta P = \left[0.171 + \frac{92.7}{Re_g}\right]\left[(\rho_g)\frac{\left(U_{ge}^2\right)}{S}\right]\left[\frac{1}{1 + C_o\,Fr_l^{0.05}}\right]^5 \tag{20.107}$$

$$Re_g = \frac{S U_{ge} \rho_g}{\mu_g} \tag{20.108}$$

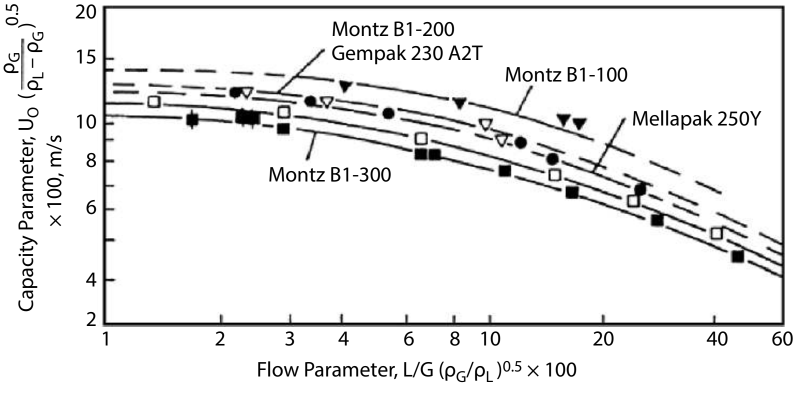


Figure 20.54 Flooding data for structured packings as reported by Billet [11]. Numbers following packing type indicate specific surface area in m^2/m^3. Reproduced by permission of the American Institute of Chemical Engineers, Fair, J. R., and Bravo, J. L., Chemical Engineering Progress, V. 86, No. (1990) p. 19. All rights reserved. Note, U_o = vapor velocity, m/s.

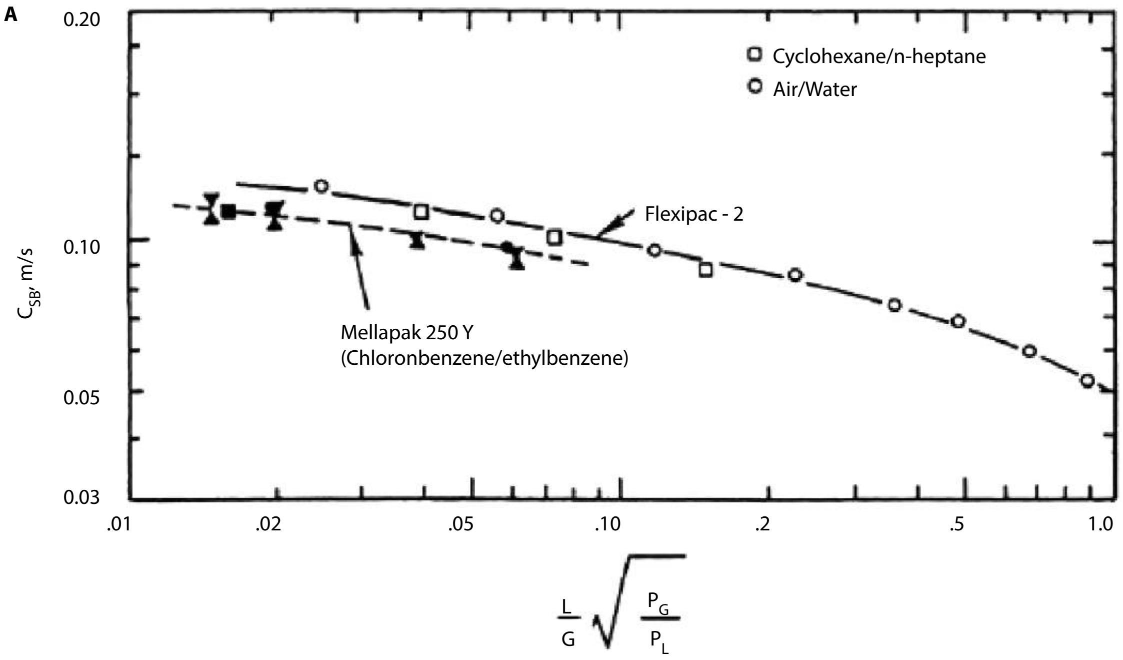


Figure 20.55A Flooding data for structured packings obtained by pressure drop measurements as well as by efficiency measurements. Reproduced by permission of the American Institute of Chemical Engineers, Fair, J. R., and Bravo, J. L., Chemical Engineering Progress, V. 86, No. 1 (1990) p. 19. All rights reserved.

$$U_{ge} = Ug/\varepsilon \sin\theta \tag{20.109}$$

$$Fr_l = U_l^2/(S_g) \tag{20.110}$$

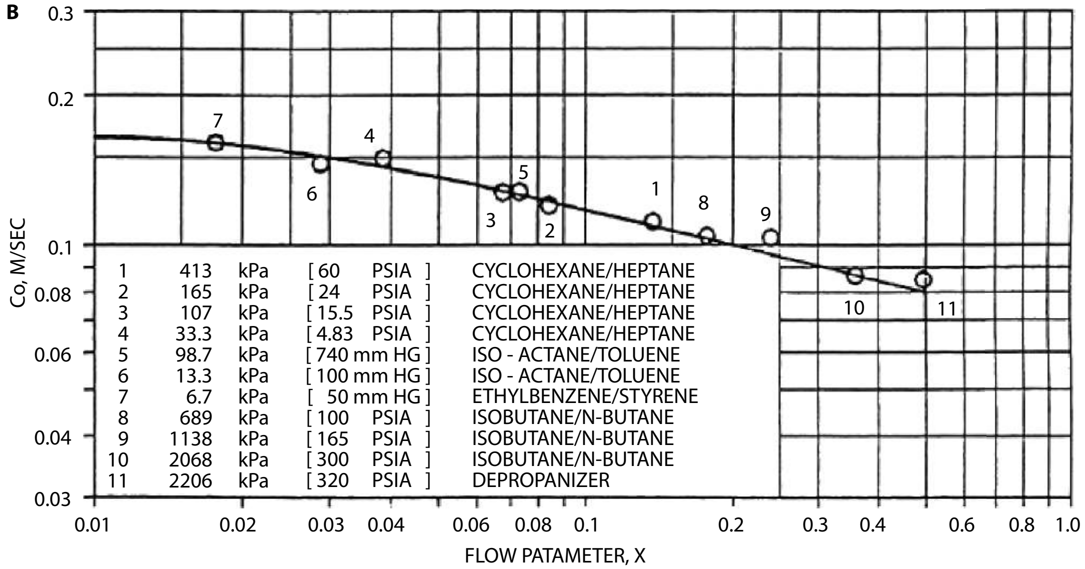


Figure 20.55B Capacity correlation of structured Intalox® 2T. The capacity parameter is at the maximum efficient capacity (MEC) point defined by Rukovena and Koshy [119]. The MEC point is lower than the flood point and located where the packing will maintain its desired efficiency. Used by permission of Rukovena, E. and Koshy, T. D., Ind. and Eng. Chem. Res. V. 32, No. 10 (1993) p. 2400. All rights reserved.

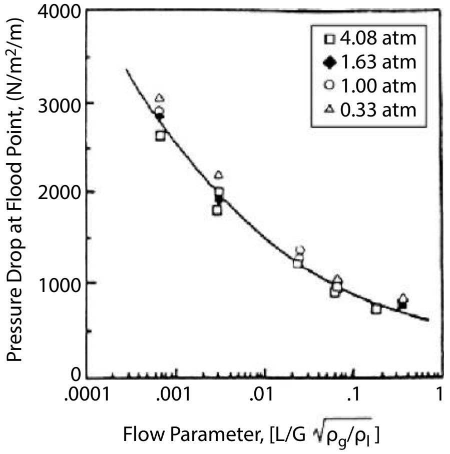


Figure 20.56 Pressure drop at the flood point as a function of loading. Values are calculated using the Stichlmair et al. model and distillation condition using cyclohexane/n-heptane and Gempak® 2A packing. Note: ordinate $(N/m^2/m)/3.385^3$ = in. Hg/m. Reproduced by permission of the American Institute of Chemical Engineers, Fair, J. R. and Bravo, J. L., Chemical Engineering Progress, V. 86, No. 1 (1990) p. 19. All rights reserved.

where
C_o = constant, value of 3.08 in. Equation 20.107 recommended for all structured packings similar to such types as Flexipac 2® and Gempak 2®. Note: Reference 72 provides values for other structured packing sizes and styles and wire gauze. Refer to manufacturer for confirmation.

where c, C_o, C_1, C_2, C_3 = correlation constants
Fr_1 = liquid Froude number
g = gravitational constant
Re_g = Reynolds number for gas
S = length of corrugation side
U_{ge} = effective velocity of gas
U_g = superficial velocity of gas
U_1 = superficial velocity of liquid
Δp = pressure drop per unit packed height
ε = packing void fraction
θ = angle of flow channel (from horizontal)
μ = viscosity
ρ = density
f (subscript) = flooding conditions
g (subscript) = gas
1 (subscript) = liquid

Billet [11] reports that metal gauze-type packing (such as Goodloe®) gives better performance and lower costs for high vacuum distillations, particularly for thermally unstable mixtures. This comparison was made against Pall rings. For the performance information referenced here, see Figure 20.57 [11].

For general references during vacuum operations:

1. Due to the rapid decrease in specific efficiency with increasing load, the optimum load factor is equal to a vapor capacity factor that in the vacuum systems is less than 60%–70% of the corresponding flood point (the actual decrease depending on the mixture). The average optimum load factor is given [11] by

$$\left(u_v \sqrt{\rho_v}\right)_{opt} = 2.4(m^{1/2}\,sec^{-1}\,kg^{1/2}) \qquad (20.111)$$

2. Specific efficiency of packing:
   $n_t/H = 5$, theoretical trays/meter
3. Average pressure drop per theoretical tray:

$$(\Delta p/n_t)_{opt} = 5 \text{ mm water column}$$

where

H = height of packing, m
n_t = number of theoretical trays
Δp = pressure drop, mm water column
u_V = vapor velocity per unit of free column cross-section, m/s.
ρ_v = density of vapor, kg/m^3

20.28.1 Guidelines for Structured Packings

1. Pressure drop: Range usually 0.3–0.5 in. water/ft, or 30–40 mm water/m of packed height [78].
2. Lower holdup of liquid: Less than other packings; may require special attention to operational controls.
3. More difficult to purge: Operating vapors more difficult to purge from the system than random packings and trays. This poses potential fire, explosion, and toxic hazard.
4. Lower pressure drop for vacuum systems: Allows for better low pressure and vacuum system operation and lower bottoms temperature, with less degradation of bottoms product.
5. Internal uniform distribution: Properly designed devices to distribute and redistribute liquid entering the column are critical to obtain best performance of these types of packings.

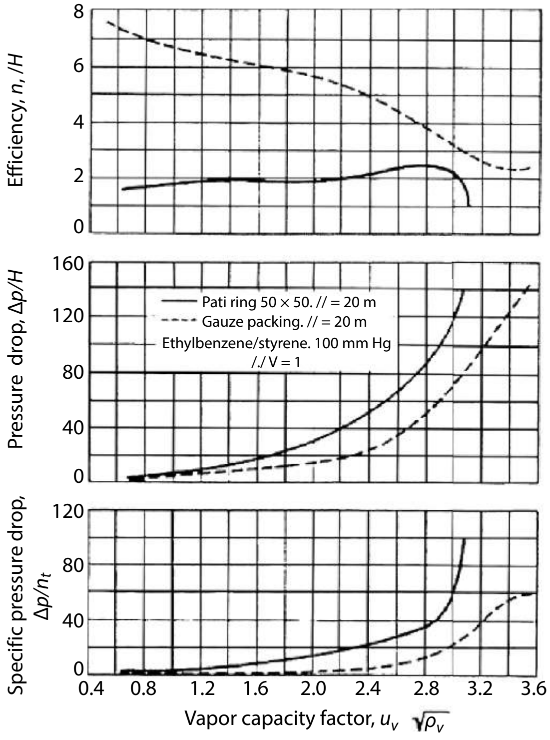


Figure 20.57 Performance of structured gauze packing vs. Pall rings. Used by permission of Billet, R., Chemical Engineering, V. 79, No. 4 (1972) p. 68. All rights reserved.

6. **Materials of construction:** The materials of fabrication for this type of packing are more critical to long life due to pad size, wall thickness of metal or plastic components, and actual selection based on the system corrosion, hydrogen attack, and oxygen attack in the column environment.
7. **HETP values:** Small packing with narrow corrugations gives low HETP values, but usually higher pressure drops. Plastic and some metal packing sheets, wires or corrugations may require special surface treatment to ensure good wettability.
8. **Preliminary designs:** These can usually be made from the generalized literature references; however, they are not a substitute for direct design of the specific system in question by the packing manufacturer, referencing to a specific packing size and style. Several competitive designs are often helpful for a final design selection.

The test studies of several different structured packings of Dean *et al.* [79] indicated that structured packing performs well in water removed by triethylene glycol (TEG) from natural gas from field wells, giving a design point of $F_s = 3$ (where $F_s = V_s\sqrt{\rho_g}$, V_s = superficial velocity, ρ_g = gas density) for sizing. A high efficiency drip-point distributor is recommended, because it was far superior to a notched trough or spray nozzle distributor. Excellent turn-down to 12:1 was shown, still meeting outlet gas low vapor specifications. HETP values ranged from 44 to 102 in. in this application. The number of theoretical trays was 2.8 to 4.6, varying with liquid and gas rates and packing style/size. Flooding occurred at 2.28 to 3.98 F_s value at 0.7 gpm/ft^2 at 650 psig.

A commercial 4-ft diameter refinery depropanizer unit's performance after replacement of the lower half of a trayed tower with Intalox-2T® structured packing is described in Reference 80.

Nutter [81] Montz B-1 structured packing, licensed from Julius Montz GmbH of Hilden, Germany, Figures 20.6LL and 20.6MM, is of uniform sinusoidal corrugations, which avoids sharp corners. The embossing on both sides of the sheet metal is closely spaced projections in a dot-matrix pattern, which spreads the liquid in a uniform film [81], without holes or slots. The design allows for an effective liquid-vapor seal at the tower wall by means of a metal wiper band. A representative performance is shown for pressure drop, Figure 20.58, and HETP, Figure 20.59. This packing is effectively used in low-pressure (or vacuum) columns where large theoretical stages and low-pressure drop are required. Style B-1 is embossed sheet metal; Style BSH is expanded textured sheet metal, high efficiency with maximum surface area; and Style A-3 is wire gauze construction.

20.28.2 Structured Packing Scale-Up

Applications of structured packing into ethylene plant's various column systems [82] have been successfully achieved, but the individual manufacturers must be consulted to use their most directly applicable pilot and commercial data, which are generally not published. The use of published general correlations should only be used for a "first"

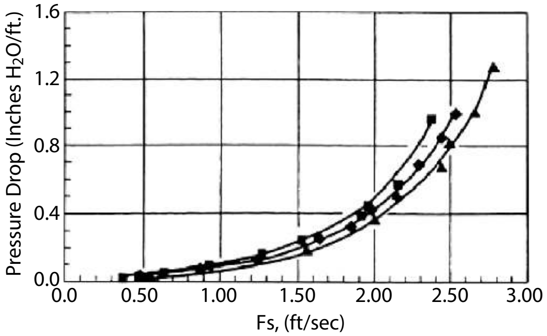


Figure 20.58 Representative pressure drop for Montz high-efficiency structured packings (several designs/styles) for a sheet metal packing. Performance can be accurately calculated by the manufacturer (See Figures 20.6LL and 20.6MM). Used by permission of Nutter Engineering Co., Harsco Corp. Bull. B-1. Specifications and descriptions used were in effect when this publication was approved for printing. Nutter Engineering reserves the right to discontinue models or options at any time, or change specifications, equipment or designs without notice and without incurring obligation. Using any of this information for specific applications should be done in consultation with Nutter Engineering personnel.

or approximation design, while the delicate or important final design must be performed in cooperation with the manufacturer.

Hufton *et al.* [83] tested gauze type packing for large commercial units, which resulted in a scale-up procedure that is too involved to reproduce here.

1. Estimate required packing height from Bravo *et al.* [75] or manufacturer's information.
2. Calculate commercial scale HETP from Fair *et al.* [72] and/or Hufton *et al.* [83].

20.29 New Generalized Pressure Drop Correlation Charts

Many generalized pressure drop correlation charts (GPDCs) have been developed over the years for random and structured type packings as shown in Figures 20.13C–I. Recently, Kister *et al.* [84] developed GPDCs for new structured and random-type packings. These charts are based upon the realization that liquid drainage in structured packings was restricted at the element to element transition rather than within each element. This implies that the liquid accumulation, which leads to flood, is initiated at the element transition region. A fourth generation of structured packings was developed in which the main body of each element has layers inclined at 45°, but the top or both ends of each element are rounded or vertical to promote drainage at this end region. These high-capacity structured packings were observed to offer more capacity than equivalent 45° inclined packings. Figures 20.60–20.72 show charts of the recent GPDC. See Kister *et al.* [84] for further details.

20.30 Mass and Heat Transfer in Packed Tower

Most packed towers are used for mass transfer operations such as absorption, distillation, and stripping; however, there are other uses, such as heat transfer quenching and entrainment knockout. The standard packings and auxiliary features associated with these towers have been presented in connection with pressure drop consideration.

Because the packed tower is a continuous contacting device, in contrast to the step-wise plate tower, performance capacity is expressed as the number of transfer units, N, the height of the transfer unit, H.T.U., and mass transfer coefficients, K_{Ga} and K_{La}. Figure 20.73 identifies the key symbols and constant flow material balance.

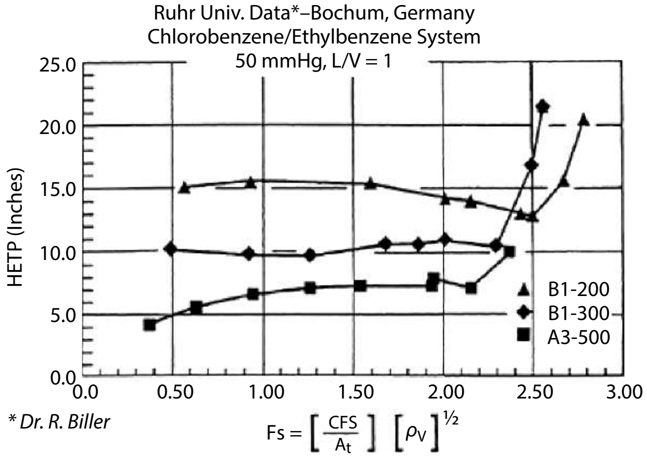


Figure 20.59 Representative HETP for Montz high-efficiency structured packings (several designs/styles) for a sheet metal packing. Used by permission of Nutter Engineering Co., Harsco Corp. Bull. B-1. Specifications and descriptions used were in effect when this publication was approved for printing. Nutter Engineering reserves the right to discontinue models or options at any time, or change specifications, equipment or designs without notice and without incurring obligation. Using any of this information for specific applications should be done in consultation with Nutter Engineering personnel.

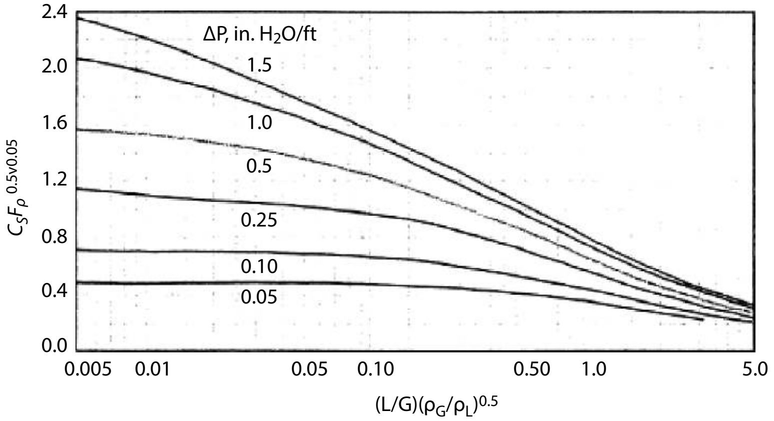


Figure 20.60 Strigle's GPDC chart for random packings. Copyright® Gulf Publishing, Houston, TX (1994).

20.31 Number of Transfer Units, N_{OG}, N_{OL}

The transfer of mass between phases in a packed tower occurs either as essentially all from gas film controlling, all from liquid film controlling, or some combination of these mechanisms (see Figure 20.74). To express the ease (low number of transfer units) or difficulty of the transfer under the conditions of operation with respect to system equilibrium, the number of transfer units N_{OG} or N_{OL} required for the system is evaluated. These can be determined experimentally and the data used for similar systems. However, it is also important to be able to estimate the number of transfer units for systems where data are not available.

$$N_{OG} = \frac{Z}{H_{OG}} \text{ or } N_{OL} = \frac{Z}{H_{OL}} \tag{20.112}$$

where

N_{OG} = number of transfer units, based on overall gas film coefficients.
N_{OL} = number of transfer units, based on overall liquid film coefficients
Z = height of packing, ft.
H_{OG} = height of transfer units, based on overall gas film coefficients, ft.
H_{OL} = height of transfer units, based on overall liquid film coefficients, ft.

The transfer process is termed gas film controlling if essentially all of the resistance to mass transfer is in the gas film. This means that the gas is usually quite soluble in, or reactive with, the liquid of the system. If the system is liquid film controlling, the gas is relatively insoluble in the liquid and the resistance to transfer is in the liquid film.

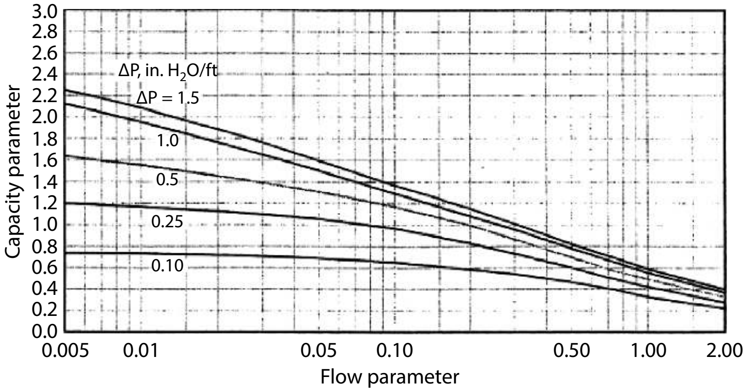
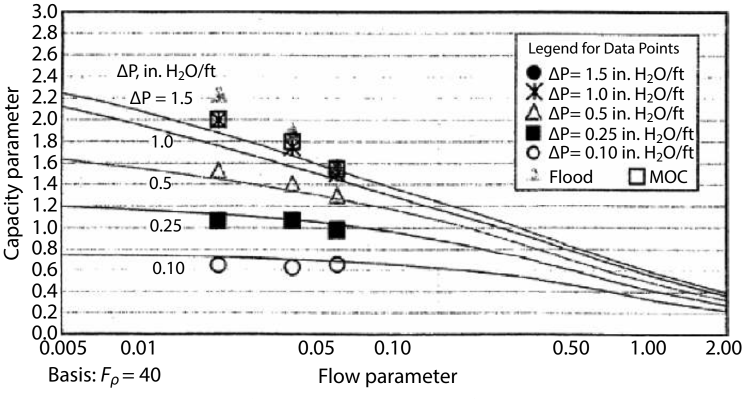
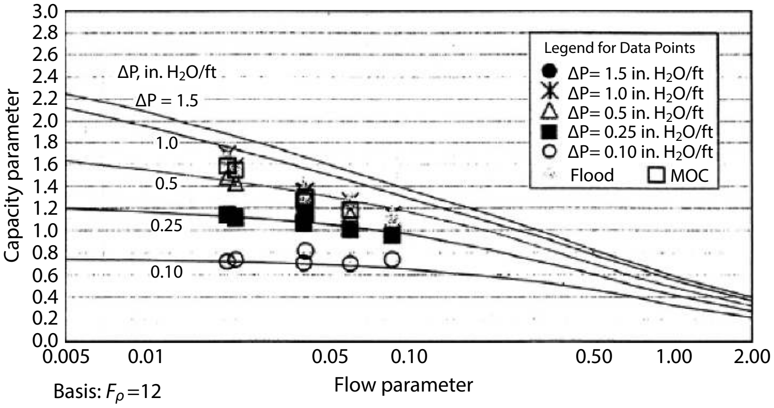


Figure 20.61 Kister and Gill's GPDC (SP) chart for structured packing. Kister *et al*. CEP, p. 28 (July 2007). By permission of AIChE. All rights reserved.



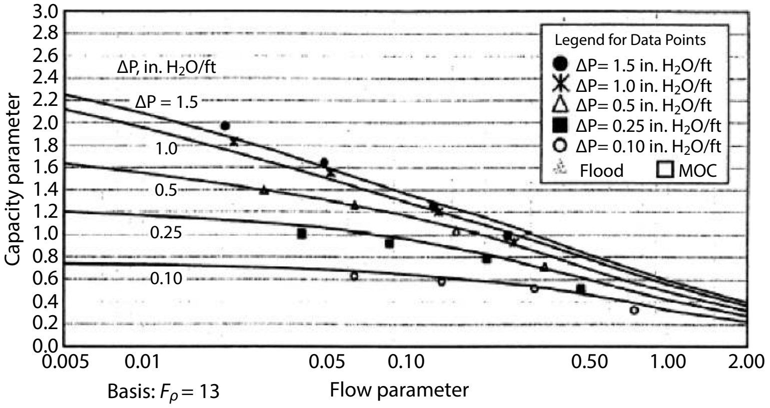
Note: The GPDC curves are for guidance only. Interpolation of the data points must be used to predict packing perfomance.

Figure 20.62 GPDC interpolation chart for Mellapak Plus 752Y with flood and pressure drop data. Kister *et al.*, CEP, p. 28, (July 2007), by permission of AIChE. All rights reserved.



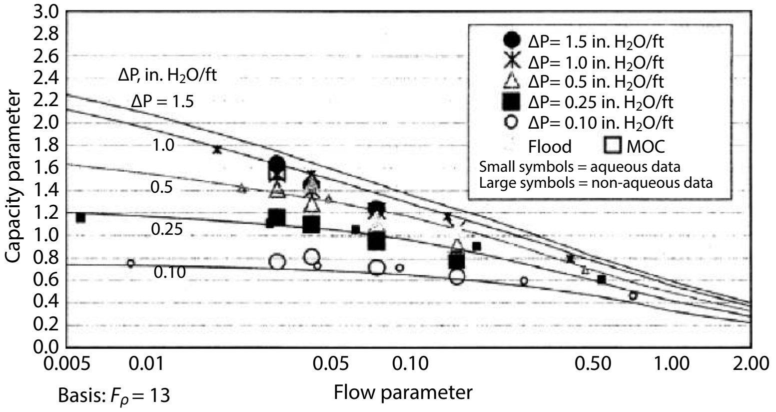
NOTE: The GPDC curves are for guidance only. Interpolation of the data points must be used to predict packing perfomance.

Figure 20.63 GPDC interpolation chart for Mellapak Plus 252Y with flood and pressure drop data. Kister *et al.*, CEP, p. 28, (July 2007), by permission of AIChE. All rights reserved.



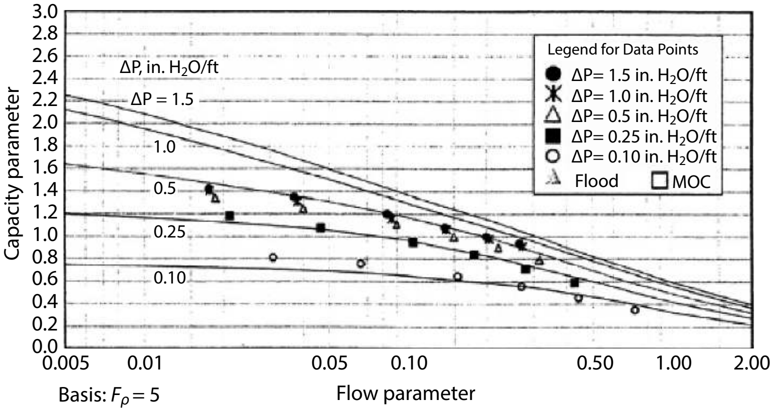
Note: The GPDC curves are for guidance only. Interpolation of the data points must be used to predict packing perfomance.

Figure 20.64 GPDC interpolation chart for Flexipac HC 2Y with pressure drop data. Kister *et al.*, CEP, p. 28, (July 2007), by permission of AIChE. All rights reserved.



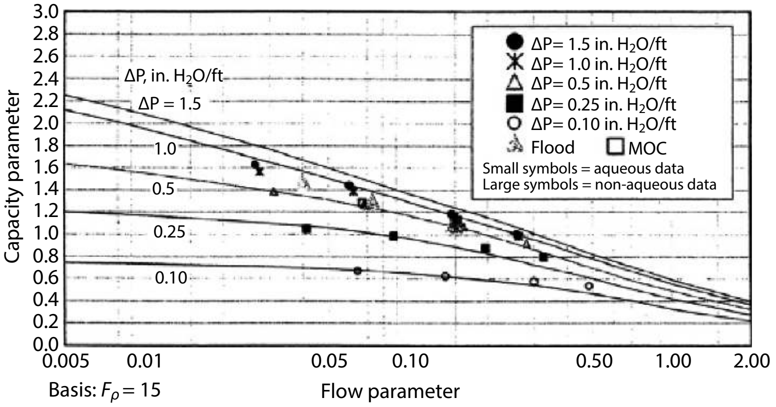
Note: The GPDC curves are for guidance only. Interpolation of the data points must be used to predict packing perfomance.

Figure 20.65 GPDC interpolation chart for Montz B1-250M with flood and pressure drop data. Kister *et al.*, CEP, p. 28, (July 2007), by permission of AIChE. All rights reserved.



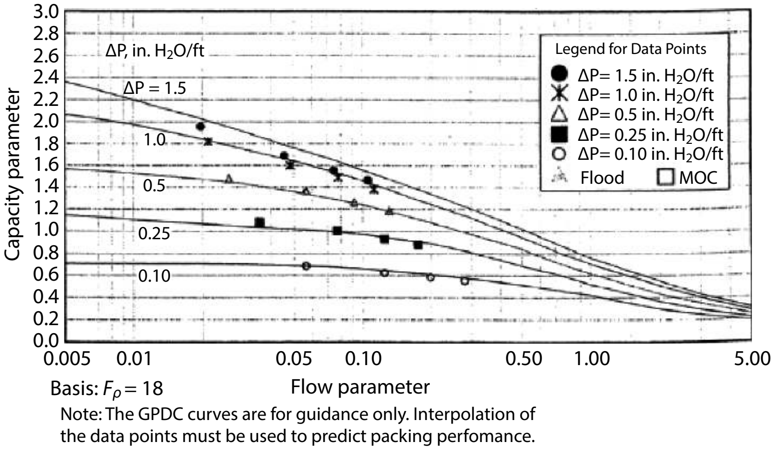
Note: The GPDC curves are for guidance only. Interpolation of the data points must be used to predict packing perfomance.

Figure 20.66 GPDC interpolation chart for Flexipac 3X with pressure drop data. Kister *et al.*, CEP, p. 28, (July 2007), by permission of AIChE. All rights reserved.



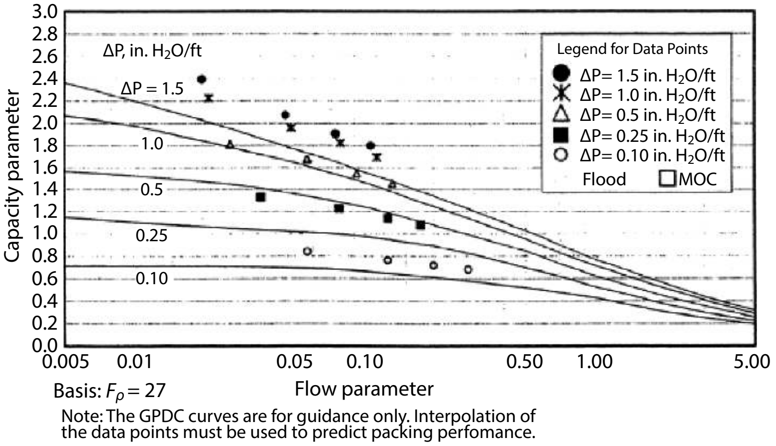
Note: The GPDC curves are for guidance only. Interpolation of the data points must be used to predict packing perfomance.

Figure 20.67 GPDC interpolation chart for Flexipac 2Y with flood and pressure drop data. Kister *et al.*, CEP, p. 28, (July 2007), by permission of AIChE. All rights reserved.



Note: The GPDC curves are for guidance only. Interpolation of the data points must be used to predict packing perfomance.

Figure 20.68 GPDC interpolation chart for # 1.5(M) Raschig Super – Ring with pressure drop data. Kister *et al.*, CEP, p. 28, (July 2007), by permission of AIChE. All rights reserved.



Note: The GPDC curves are for guidance only. Interpolation of the data points must be used to predict packing perfomance.

Figure 20.69 GPDC interpolation chart for # 1.5(M) Raschig Super-Ring with a grossly excessive packing factor. Kister *et al.*, CEP, p. 28, (July 2007), by permission of AIChE. All rights reserved.

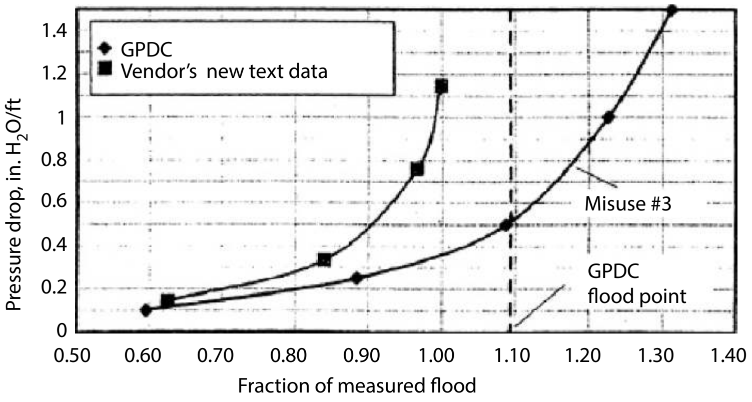


Figure 20.70 Incorrect application of GPDC beyond the flood point. Kister *et al.*, CEP, p. 28, (July 2007), by permission of AIChE. All rights reserved.

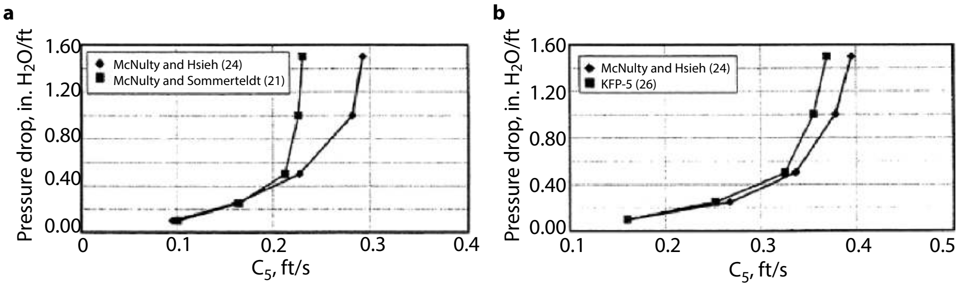
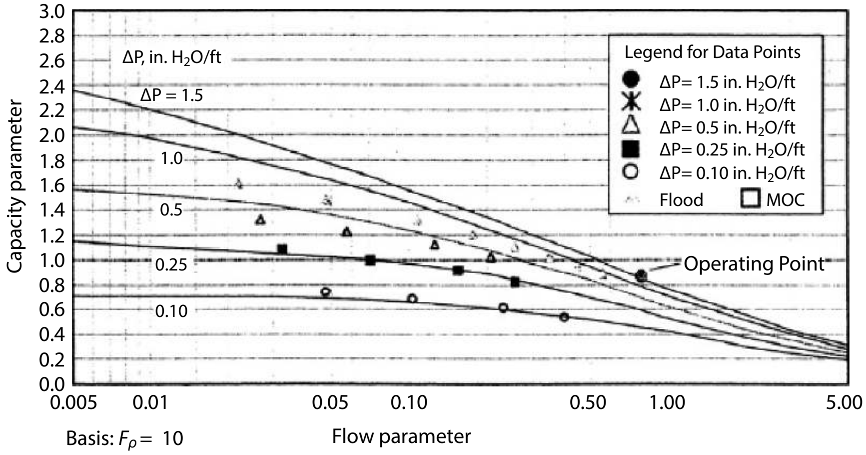


Figure 20.71 (a) Flexipac 1Y metal structured packing pressure drop vs. C-factor (air–water, 10 gpm/ft², Koch/Koch-Giltsch data). Kister *et al.*, CEP, p. 28, (July 2007), by permission of AIChE. All rights reserved. (b) Flexipac 2Y metal structured packing pressure drop vs. C-factor (air–water, 10 gpm/ft², Koch/Koch-Giltsch data). Kister *et al.*, CEP, p. 28, (July 2007), by permission of AIChE. All rights reserved.



Note: The GPDC curves are for guidance only. Interpolation of the data points must be used to predict packing perfomance.

Figure 20.72 GPDC interpolation chart for packing in tower stripping section, showing an operating point above flood. Kister *et al.*, CEP, p. 28, (July 2007), by permission of AIChE. All rights reserved.

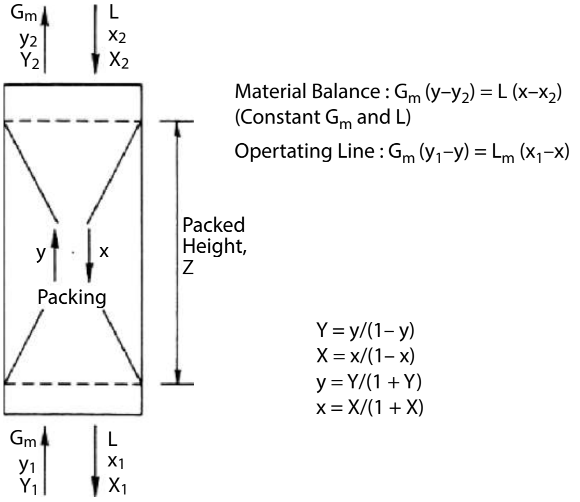


Figure 20.73 Counter current packed tower symbols.

Many systems are a combination of the two. Without good data, it is next to impossible to design equipment exactly, although satisfactory designs are possible. System information is presented in Table 20.41 as a guide. Other data for different systems exist in the literature in a scattered fashion.

For (1) dilute solutions or (2) equal molar diffusion between phases (e.g., distillation)

$$N_{OG} = \int_{y_2}^{y_1} \frac{dy}{(y-y^*)} = \frac{y_1 - y_2}{\dfrac{(y-y^*)_1 - (y-y^*)_2}{\ln\dfrac{(y-y^*)_1}{(y-y^*)_2}}} \tag{20.113}$$

Table 20.41 System film control*.

Gas Film
1. Absorption of ammonia in water
2. Absorption of ammonia in aqueous ammonia
3. Stripping of ammonia from aqueous ammonia
4. Absorption of water vapor in strong acids
5. Absorption of sulfur trioxide in strong sulfuric acid
6. Absorption of hydrogen chloride in water
7. Absorption of hydrogen chloride in weak hydrochloric acid
8. Absorption of 5 vol. percent ammonia in acids
9. Absorption of sulfur dioxide in alkali solutions
10. Absorption of sulfur dioxide in ammonia solutions
11. Absorption of hydrogen sulfide in weak caustic
12. Evaporation of liquids
13. Condensation of liquids
Liquid film
1. Absorption of carbon dioxide in water
2. Absorption of oxygen in water
3. Absorption of hydrogen in water
4. Absorption of carbon dioxide in weak alkali
5. Absorption of chlorine in water
Both gas and liquid film
1. Absorption of sulfur dioxide in water
2. Absorption of acetone in water
3. Absorption of nitrogen oxide in strong sulfuric acid

*From M. Leva, *Tower Packings and Packed Tower Design*, 2nd Ed. p. 91. U.S. Stoneware Co. (1953). By permission, now, Norton Chemical Process Products Corp.

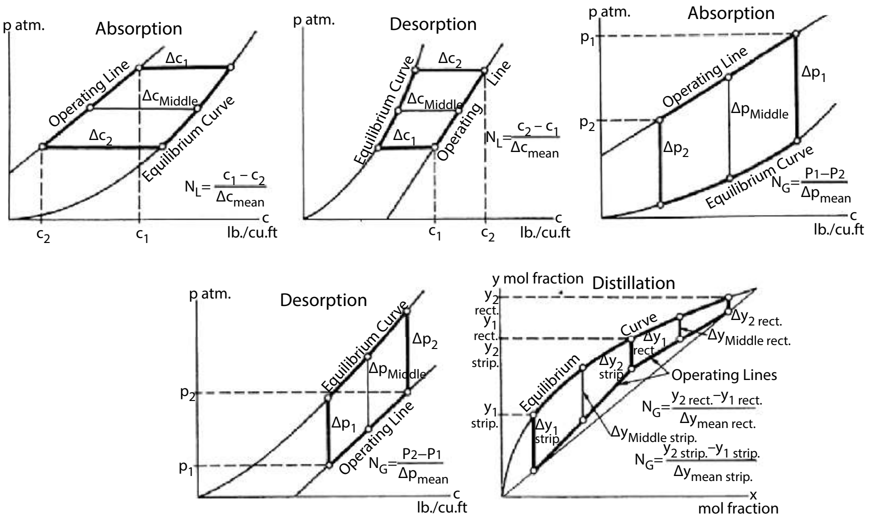


Figure 20.74 Mass transfer diagrams. The number of transfer units can be determined by the difference in concentration or vapor pressure, particularly over ranges where the equilibrium line is essentially straight. Used by permission of Czermann, J. J., Gyokheqyi, S. L., and Hay, J. J., Petroleum Refiner, V. 37, No. 4 (1958) p. 165. All rights reserved.

where $(y - y^*)$ = driving force, expressed as mol fractions
y = mol fraction of one component (solute) at any point in the gas phase of the contacting system.
y^* = mol fraction gas phase composition in equilibrium with a liquid composition, x
x = mol fraction in the liquid at the same corresponding point in the system as y.
1, 2 = inlet and outlet conditions of system.

If the system has more than two components, the calculations may be based on the component that varies the most in passing through the unit, or the component for which good data are available.

A large majority of systems have operating lines and equilibrium curves that can be assumed to be straight over the range covered by the design. For the conditions of a straight-line equilibrium curve, $y^* = mx$, Colburn [85, 86] has integrated the relation above to obtain

$$N = \frac{2.3 \log[(1 - P'')M + P'']}{1 - P''} \tag{20.114}$$

where N may be N_{OG} or N_{OL} depending on the operation.

Table 20.42 identifies several important conditions that affect the values of P″ and M. These are extracted from Colburn's larger summary [86].

Figure 20.75 is a plot to aid in solving the equation for N (or N_{OG} or N_{OL}).

For constant temperature absorption, with no solute in the inlet liquid, $x_2 = 0$, and the abscissa becomes y_1/y_2.

For Concentrated Solutions and More General Application

The following equation applies for diffusion in one direction (e.g., absorption, extraction, desorption) [88]:

$$N_{OG} = \int_{y_2}^{y_1} \frac{(1-y)_M \, dy}{(1-y)(y-y^*)} \tag{20.115}$$

$$N_{OG} = \int_{y_2}^{y_1} \frac{dy}{(y - y^*)} + \frac{1}{2}\ln\frac{(1 - y_2)}{(1 - y_1)} \tag{20.116}$$

or

$$N_{OG} = \int_{Y_2}^{Y_1} \frac{dY}{(Y - Y^*)} - \frac{1}{2}\ln\frac{(1 + Y_1)}{(1 + Y_2)} \tag{20.117}$$

Table 20.42 Values to use with Transfer Equation and Figure 20.75.

Condition of operation	**p''**	**M**
Absorption:		
1. Constant mG_m/L_m	mG_m/L_m	$(y_1 - mx_2) / (y_2 - mx_2)$
2. Varying mG_m/L_m	m_2G_{m2}/L_{m2}	$\left(\frac{(y_1 - m_2x_2)}{(y_2 - m_2x_2)}\right)\left(\frac{1 - m_2G_m/L_m}{1 - y_1^*/y_1}\right)$
Desorption (stripping):		
3. Constant L_m/mG_m	L_m/mG_m	$(x_1 - y_2/m) / (x_2 - y_2m)$
4. Varying L_m/mG_m	L_{m2}/m_2G_{m2}	$\left(\frac{x_1 - y_2/m_2}{x_2 - y_2/m_2}\right)\left(\frac{1 - L_{m2}/m_2G_{m2}}{1 - x_1^*/x_1}\right)$
Distillation, enriching[1]		
5. Constant mG_m/L_m		same as 1
6. Varying mG_m/L_m stripping, closed steam[2]		same as 2
7. Constant L_m/mG_m	L_m/mG_m	$(x_1 - x_2/m)/(x_2 - x_2/m)$
8. Varying L_m/mG_m stripping, open steam[2]	L_{m2}/mG_{m2}	$\left(\frac{x_1 - x_2/m_2}{x_2 - x_2m_2}\right)\left(\frac{1 - L_2/m_2G_{m2}}{1 - x_1^*/x_1}\right)$
9. Constant L_m/mG_m	L_m/mG_m	x_1/x_2
10. Varying L_m/mG_m	L_{m2}/m_2G_{m2}	$\left(\frac{x_1}{x_2}\right)\left(\frac{1 - L_{m2}/m_2G_{m2}}{1 - x_1^*/x_1}\right)$

*Equilibrium value

Subscripts 1 and 2 refer to the concentrated and dilute ends of the unit, respectively

[1] Concentrations and m are based on high boiler or "heavy key"

[2] Concentrations and m are based on low boiler or "light key"

m = slope of equilibrium line (mol-fraction solute in gas)(mol-fraction solute in liquid)

By permission, A. P. Colburn, *Ind. Eng. Chem.* 33, 459 (1941). The American Chem. Soc., all rights reserved.

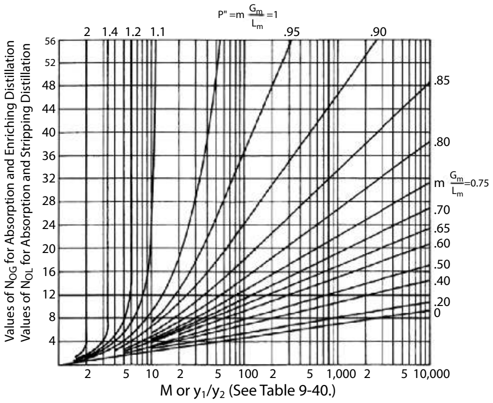


Figure 20.75 Colburn plot for transfer units. Reproduced by permission of the American Chemical Society; Colburn, A. P., Ind. Chem. V. 33 (1941) p. 459. All rights reserved.

or

$$N_{OG} = \int_{Y_2}^{Y_1} \frac{dY}{(Y - Y^*)} - 1.15 \log_{10} \frac{(1 + Y_1)}{(1 + Y_2)} \tag{20.118}$$

where $(1 - y)_M$ = log mean average of concentration at the opposite ends of the diffusion process, (1 – y) in the main gas body, and (1 – y*) at the interface [88].

y = Concentration of solute in gas, mol fraction.
y* = Concentration of solute in gas in equilibrium with liquid, mol fraction.
Y = Concentration of solute in gas, lb mol solute/lb mol solvent gas.
Y* = Concentration of solute in gas in equilibrium with liquid, lb mol solute/lb mol solvent gas.

If the liquid film controls:

$$N_{OL} = \int_{x_2}^{x_1} \frac{dx}{(x^* - x)} + \frac{1}{2} \ln \frac{(1 - x_2)}{(1 - x_1)} \tag{20.119}$$

$$N_{OL} = \int_{X_2}^{X_1} \frac{dX}{(X^* - X)} - \frac{1}{2} \ln \frac{(1 + X_1)}{(1 - X_2)} \tag{20.120}$$

where

x = Concentration of solute in liquid, mol fraction.
x* = Concentration of solute in liquid in equilibrium with gas, mol fraction.
X = Concentration of solute in liquid, lb mol solute/lb mol solvent.
X* = Concentration of solute in liquid in equilibrium with gas, lb mol solute/lb mol solvent.

It is usually necessary to graphically integrate the first terms of the above equations, although some problems do allow for mathematical treatment.

Example 20.10

A gas stream comprising 92 mole % dry air and 8 mole % SO_2 is to be scrubbed with fresh water in a packed tower. The feed flow rate will be 60 kmol/h of gas and the exit gas should contain less than 0.18% SO_2.

The tower is to be designed to operate at 70% of flooding velocity. Operation will be isothermal at 30°C and 1 bar. The water flow rate is to be twice the minimum flow rate.

Calculate:

The tower diameter.

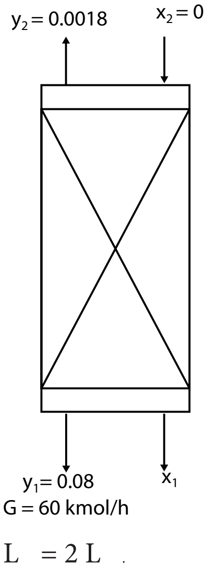
The packed height required, assuming the height of an overall gas phase transfer unit is 0.9 m.

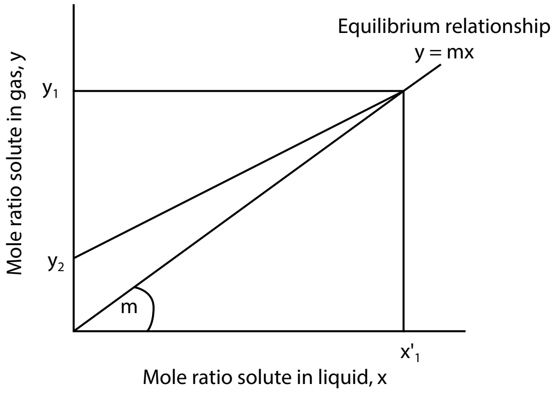
Data:

Equilibrium relationship:	$y^* = 25.4x$
Packing factor	$a_p = 253\ m^2/m^3$
Voidage	$e = 0.79$
Density of water	$\rho_L = 1000\ kg/m^3$
Acceleration due to gravity	$g = 9.81\ m/s^2$
Molar mass of air	29.0 kg/kmol
Molar mass of SO_2	64 kg/kmol

1 kmol of gas at 0°C and 1 bar occupies 22.4 m^3

Solution:





$L = 2\,L_{min}$
$u = 70\ \%\ u_{Flood}$
$y^* = 25.4\,x$
$e = 0.79$
$a_p = 253$
$G' = (1 - 0.08) \times 60$
$= 55.2$ kmol/h

For minimum flow:

$$0.08 = 25.4\,x_1'$$

$$x_1' = 0.0032$$

Operating Line Equation:

$$G'\left[\frac{y_1}{1-y_1} - \frac{y_2}{1-y_2}\right] = L'_{min}\left[\frac{x_1}{1-x_1} - \frac{x_2}{1-x_2}\right]$$

$$55.2\left[\frac{0.08}{0.92} - \frac{0.0018}{0.9982}\right] = L'_{min}\left[\frac{0.0032}{0.9968} - 0\right]$$

$$55.2 \times 0.08516 = L'_{min} \times 0.00321$$

$$L'_{min} = 1464.4 \text{ kmol/h}$$

Actual liquid rate, $L = 2L'_{min} = 2928$ kmol/h

SO_2 transferred = 60 – 55.2 = 4.8 kmol/h.

$$x_1 = \frac{4.8}{2928 + 4.8} = 0.0016$$

Total flow rate:

G_{in} = 55.2 × 29 + 4.8 × 64 = 1908 kg/h
G_{out} = 55.2 × 29 + 0 = 1600.8 kg/h
L_{in} = 2928 × 18 + 0 = 52704 kg/h
L_{out} = 52704 + 4.8 × 64 = 53011.2 kg/h

Densities:

Average molecular weight, Mw_{avg} = 0.08 x 64 + 0.92 x 29 = 31.8 kg/kmol
Density of water, ρ_L = 1000 kg/m³ Liquid:
Density of gas, ρ_G = 1 kmol at NTP occupies 22.4 m³

$$31.8 \text{ kg at } 25^{\circ}\text{C occupies } 22.4 \times \frac{298}{273} = 24.45\,\text{m}^3$$

Therefore, density of gas is: $\rho_G = \frac{31.8}{24.45} = 1.3006 \frac{\text{kg}}{\text{m}^3}$

$$\frac{L}{G}\sqrt{\frac{\rho_G}{\rho_L}} = \frac{52704}{1908}\sqrt{\frac{1.3006}{1000}} = 0.996$$

From Figure 20.76, the intercept = 0.019

$$0.019 = \frac{u_F^2}{g}\frac{a_p}{e^3}\left(\frac{\rho_G}{\rho_L}\right)\left(\frac{\mu_L}{\mu_W}\right)^{0.2}$$

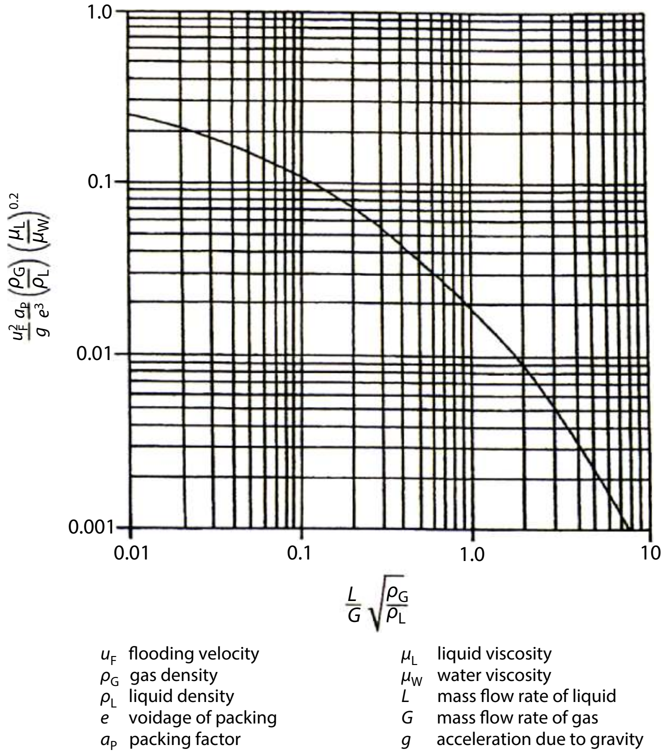


Figure 20.76 Generalized correlation for flooding rates in packed towers for Example 20.10.

where $\mu_L = \mu_W$

$$0.019 = \frac{u_F^2}{9.81}\frac{253}{0.29^3}\left(\frac{1.3006}{1000}\right)$$

$$u_F = 0.5288 \text{ m/s}$$

The gas velocity at 70% of flooding velocity is: u = 0.7 x 0.5288 = 0.3699 m/s.

Volumetric gas flow = 60 x 24.45 (kmol/h x m^3/kmol)

$$= 146.70 \text{ m}^3/\text{h}$$

$$= 0.4075 \text{ m}^3/\text{s}$$

Area of the tower = 0.4075 / 0.3699 = 1.1016 m^2

Tower diameter from the area:

$$d = \sqrt{\frac{4\,\text{Area}}{\pi}} = \sqrt{\frac{4 \times 1.1016}{\pi}}$$
$$= 1.184 \text{ m.}$$

(ii) The number of transfer unit is

$$\mathrm{NTU_{OG}} = \frac{1}{(1-1/A)} \ln\left[\left(1-\frac{1}{A}\right)\frac{y_1}{y_2} + \frac{1}{A}\right]$$

$$A = \frac{L'}{G'm} = \frac{2928}{25.4 \times 55.2} = 2.09$$

$$\mathrm{NTU_{OG}} = \frac{1}{\left(1-\frac{1}{2.09}\right)} \ln\left[\left(1-\frac{1}{2.09}\right)\frac{0.08}{0.0018} + \frac{1}{2.09}\right]$$

$$\mathrm{NTU_{OG}} = 6.07$$

Packed height, $z = \mathrm{HTU_{OG}} \times \mathrm{NTU_{OG}}$
$z = 0.9 \text{ x } 6.07 = 5.46$ m.

20.32 Gas and Liquid-Phase Coefficients, k_G and k_L

Recent studies indicate that individual film transfer coefficients may be correlated with good agreement for Raschig rings and Berl saddles for aqueous and non-aqueous systems [65]:

$$\begin{aligned} j_D &= \left[\frac{k_G M_M P_{BM}}{G}\right]\left[\frac{\mu_{Ga}}{\rho_G D_V}\right]^{0.667} \\ &= 1.195\left[\frac{D_p G'}{\mu_{Ga}(1-\varepsilon)}\right]^{-0.36} \end{aligned} \tag{20.121}$$

This has been shown to correlate for a wide variety of tower packings, various operating conditions, and physical properties of solute and inert gases. The calculated k_G must be used in conjunction with the effective interfacial areas determined by Shulman [64] (Figure 20.39) to establish a reliable value for k_Ga. Figure 20.39 should be used with the abscissa as $G/\sqrt{\rho/0.075}$ for inert gas other than air [65]:

$$k_Ga = k_G\,(a)$$

$$\mathrm{HTU(gas\ phase)} = G'/k_G a\, M_M\, P_{BM} \tag{20.122}$$

where
D_V = diffusivity of solute in gas, ft^2/h
k_G = gas-phase mass transfer coefficient, lb mol/h.ft^2 atm.
M_M = mean molecular weight of gas, lb/lb mol
P_{BM} = mean partial pressure of inert gas in the gas phase, atm.
μ_{Ga} = gas viscosity, lb/h.ft.

ρ_G = gas density, lb/ft^3
G′ = superficial gas rate, lb/h.ft^2

$$\frac{1}{K_G a} = \frac{1}{k_G a} + \frac{1}{H' k_L a} = \frac{1}{k_G a} + \frac{m}{k_L a} = \frac{1}{H' K_L a} \tag{20.123}$$

$$\frac{1}{K_L a} = \frac{1}{k_L a} + \frac{H'}{k_G a} = \frac{1}{K_L a} + \frac{1}{m k_G a} \tag{20.124}$$

H′ = Henry’s law constant, lb mols/(ft^3) (atm)
k_L = liquid-phase mass transfer coefficient, lb mols/h.ft^2 (lb mols/ft^3)

The relation

$$\frac{\text{vaporization}}{\text{absorption}} = \frac{(k_G a)v}{(k_G a)a} = 0.85\frac{h_t}{h_o} \tag{20.125}$$

is reported to correlate with ±8% based on test data and appears to be founded on a sound investigative program.

For the liquid phase based on Raschig ring and Berl saddle data [64]:

$$\frac{k_L D_P}{D_L} = 25.1\left[\frac{D_P L'}{\mu_{La}}\right]^{0.45}\left[\frac{\mu_{La}}{\rho_L D_L}\right]^{0.5} \tag{20.126}$$

Use of k_G and k_L

1. From physical properties of system, determine k_G and k_L. If the system is known, or can be assumed to be essentially all gas or all liquid film controlling, then only the controlling k need be calculated. For greater accuracy, both values are recommended, because very few systems are more than 80% controlled by only one k.
2. Combine effective interfacial area to calculate $k_G a$ or $k_L a$.
3. Determine $k_G a$ by:

$$\frac{1}{K_G a} = \frac{1}{k_G a} + \frac{1}{H' k_L a} = \frac{1}{H' K_L a} \tag{20.127}$$

20.33 Height of a Transfer Unit, H_{OG}, H_{OL}, HTU

The earlier concept of height equivalent to a theoretical plate (HETP), which relates the height of packing to a unit of transfer known as the theoretical stage or plate, has generally been dropped in favor of the “height of a transfer unit” HTU. This is designated as HOG or HOL depending on whether it was determined from gas or liquid film data. HETP data for absorption and distillation are given in the section under packed tower distillation.

$$\frac{\text{HETP}}{H_{OG}} = \frac{(mG_m/L_m - 1)}{\ln(mG_m/L_m)} \tag{20.128}$$

Height of Overall Transfer Unit

For small changes in concentration and if the total number mols of gas and liquid remains essentially constant, the following is applicable to all but very concentrated solutions. For the latter case, see References 88 and 89.

$$H_{OG} = \frac{G_m}{K_G a\, P_{ave}} \tag{20.129}$$

$$H_{OL} = \frac{L_m}{K_L a\, \rho_L} \tag{20.130}$$

$$(HTU)_{OG} = \frac{V'}{K_L a\, \rho_L} \tag{20.131}$$

$$Z = (HTU)_{OG}\,(NTU)_{OG} \tag{20.132}$$

Height of Individual Transfer Unit

These calculations apply to the same limiting conditions as the preceding section. Some data are reported as individual gas or liquid film coefficients or transfer unit heights; however, it is often possible to use it as overall data if the conditions are understood.

$$H_G = \frac{G_m}{k_G a\, P_{ave}} \tag{20.133}$$

$$H_L = \frac{L_m}{k_L a\, \rho_L} \tag{20.134}$$

$$H_{OG} = H_G + \frac{mG_m}{L_m}(H_L) = H_G + \frac{H_L}{A'} \tag{20.135}$$

$$H_{OL} = H_L + \frac{L_m}{mG_m}(H_G) = H_L + A' H_G \tag{20.136}$$

where $A' = L_m / mG_m$

For predominantly liquid film controlling systems, $A'H_G$ is almost negligible and $H_{OL} = H_L$; likewise for gas film controlling, H_L / A' is negligible and $H_{OG} = H_G$.

where

G_m = gas mass velocity, lb mol/h. ft^2

L_m = liquid mass velocity, lb mol/h. ft^2

$K_G a$ = overall gas mass-transfer coefficient, lb mol/(h.) (ft^3) (atm)

$K_L a$ = overall liquid mass-transfer coefficient, lb mol/(h.) (ft^3) (lb mol/ft^3)

$k_G a$ = individual gas mass-transfer coefficient, lb mol/(h.) (ft^3) (atm)

Table 20.43 Liquid film height of transfer unit*.

Packing	**φ**	**j**	**Range of L´Lb/hr (ft^2)**
Raschig Rings (in.)			
3/8	0.00182	0.46	400–15,000
½	0.00357	0.35	400–15,000
1	0.0100	0.22	400–15,000
1.5	0.0111	0.22	400–15,000
2	0.0125	0.22	400–15,000
Berl Saddles (in.)			
½	0.00666	0.28	400–15,000
1	0.00588	0.28	400–15,000
1.5	0.00625	0.28	400–15,000
3 In. Partition rings, stacked staggered	0.0625	0.09	3,00–14,000
Spiral Rings, stacked staggered			
3-in. single spiral	0.00909	0.28	400–15,000
3-in. triple spiral	0.0116	0.28	3,000–14,000
Drip-point grids (continuous flue)			
Style 6146	0.0154	0.23	3,500–30,000
Style 6295	0.00725	0.31	2,500–22,000

*Reproduced by permission, Treybal, R. E., *Mass Transfer Operations*, McGraw-Hill Book Co., Inc. (1955), p. 237, using data of Sherwood, T. K. and Holloway, F. A. L. [62] and of Molstad, McKinney and Abbey [51], all rights reserved.

k_La = individual liquid mass-transfer coefficient, lb mol/(h.) (ft^3) (lb mol/ft^3)
P_{ave} = average total pressure in tower, atm.
H_L = height of liquid film transfer unit, ft.
H_G = height of gas film transfer unit, ft.
a = effective interfacial area for contacting gas and liquid phases, ft^2/ft^3. Because this is very difficult to evaluate, it is usually retained as a part of the coefficient such as K_Ga, K_La, k_Ga, and k_La.
V′ = vapor flow rate, lb mol/h.

Estimation of Height of Liquid Film Transfer Units

The following relation is used in estimating liquid film transfer units [54]. For the proper systems, H_L may be assumed to be equal to H_{OL}.

$$H_L = \phi(L'/\mu_{La})^j(\mu_{La}/\rho_L D_L)^{0.5} \tag{20.137}$$

where $\mu_{La}/\rho_L D_L$ = Schmidt number
H_L = height of transfer unit, ft.
L' = liquid rate, lb/(h.) (ft^2)

Table 20.44 Diffusion coefficients of gases and liquid in liquids at 68°F (Dilute Concentration).

	Solvent	**Diffusion coefficient, D_L ft²/hr (multiply all values by 10^{-5})**
Gas		
Oxygen	Water	7.0
Carbon Dioxide	Water	5.82
Nitrous Oxide	Water	5.86
Ammonia	Water	6.8S
Chlorine	Water	4.74
Bromine	Water	4.66
Hydrogen	Water	19.92
Nitrogen	Water	6.37
Hydrogen Chloride	Water	10.25
Hydrogen Sulfide	Water	5.47
Acetylene	Water	6.06
Liquid		
Sulfuric Acid	Water	6.72
Nitric Acid	Water	10.15
Methanol	Water	4.97
Sodium Chloride	Water	5.23
Sodium Hydroxide	Water	5.86

Note: Additional data are given in the references, as well as the International Critical Tables.
From: Perry J. H., *Chem. Engrs. Handbk*. 3rd Ed. p. 540, McGraw-Hill Book Co., Inc. © (1950), By permission.

μ_L = viscosity of liquid, lb/ft. h
D_L = liquid diffusivity, ft²/h

ϕ and j are constants given in Table 20.43.

Diffusivity values are given in Table 20.44.

Estimation of Height of Gas Film Transfer Units

The relation [53, 54, 90]

$$H_G = \frac{\alpha G'\beta}{L^{\gamma}}\left(\frac{\mu_{Ga}}{\rho_G D_G}\right)^{0.5} \tag{20.138}$$

describes a reasonable part of the gas film data. It allows the conversion of the ammonia–air–water data of Fellinger [120] to other systems. Table 20.45 gives the constants for the equation.

α, β, γ = constants peculiar to packing for dilute and moderate concentration [88]:

$$H_{OG} = H_G + \frac{mG}{L_m}(H_L) = H_G + \frac{H_L}{A} \tag{20.139}$$

$$H_{OL} = H_L + \frac{L_m}{mG_m}(H_G) = H_L + AH_G \tag{20.140}$$

Figure 20.77 presents some of the data of Fellinger [120], as presented in Reference 31 for H_{OG} for the ammonia–air–water systems. These data may be used with the Sherwood relations to estimate H_L and H_G values for other systems.

Estimation of Diffusion Coefficients to Gases

Good, reliable diffusion data are difficult to obtain, particularly over a wide range of temperatures. The Gilliland relation is [90]

$$D_V = 0.0069\frac{T^{3/2}\sqrt{1/M_A + 1/M_B}}{P\left(V_A^{1/3} + V_B^{1/3}\right)^2}, \text{ft}^2/\text{h} \tag{20.141}$$

Table 20.45 Gas film height of transfer unit.

Packing	α	β	ϒ	G′	L′
Raschig Rings (in.)					
⅜	2.32	0.45	0.47	200–500	500–1,500
1	7.00	0.39	0.58	200–800	400–500
	6.41	0.32	0.51	200–600	500–4,500
1.5	17.3	0.38	0.66	200–700	500–1,500
	2.58	0.38	0.40	200–700	1,500–4,500
2	3.82	0.41	0.45	200–800	500–4,500
Bed Saddles (in.)					
½	32.4	0.30	0.74	200–700	500–1,500
	0.811	0.30	0.24	200–700	1,500–4,500
1	1.97	0.36	0.40	200–800	400–4,500
1.5	5.05	0.32	0.45	200–1,000	400–4,500
3-in. Partition Rings (stacked staggered)	650	0.58	1.06	150–900	3,000–10,000
Spiral Rings (stacked staggered)					
3-in. single spiral	2.38	0.35	0.29	130–700	3,000–10,000
3-in. triple spiral	15.6	0.38	0.60	200–1,000	500–3,000
Drip Point Grids (continuous flue)					
Style 6146	3.91	0.37	0.39	130–1,000	3,000–6,500
Style 6295	4.65	0.17	0.27	100–1,000	2,000–11,500

From the data of Fellinger [27] and of Molstad *et al.* [50, 51] as presented in Treybel, R. E. *Mass Transfer Operations*, p. 239, McGraw-Hill Book Co. Inc. © (1955), Ref. 74, by permission, all rights reserved.

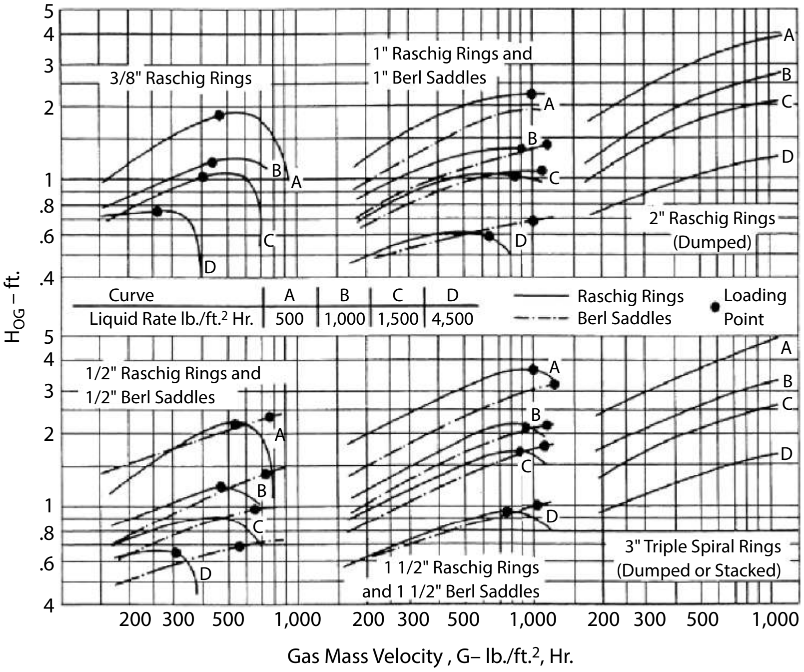


Figure 20.77 Fellinger's overall gas film mass transfer data for ammonia – water system. Used by permission of Leva, M. Tower Packings and Packed Tower Design, 2nd. Ed. U.S. Stoneware Co. (now, Norton Chemical Process Products Corp) (1953).

where

T = absolute temperature, °R
M_A, M_B = molecular weights of the two gases, A and B
P = total pressure, atm
V_A, V_B = molecular volumes of gases, obtained by Kopp's law of additive volume, cm³/gm mol at normal boiling point. See Table 20.46.

Diffusion coefficients are used to estimate $K_G a$ values for gas film controlling systems:

K_G a (unknown systems) = K_G a (known system)

$$\left[\frac{D_V \text{ unknown}}{D_V \text{ known}}\right]^{0.56} \tag{20.142}$$

Ammonia–air–water system data, Figure 20.77, is often used by converting H_{OG} (ammonia-air) to its corresponding $K_G a$, and then substituting the above relation for the unknown $K_G a$.

20.34 Distillation in Packed Towers

Packed towers are used in some distillation operations in preference to plate towers. Usually the selection requires an understanding of the fouling characteristics of the fluids of the system. These towers even have been used in polymer forming operations. However, other contacting devices can be cleaned more easily. For some processes, the packed tower is much more effective as well as cheaper than a tray tower.

The more complicated separation for a three-phase system is discussed by Harrison [91].

Table 20.46 Atomic and molecular volumes.

Atomic volume		Molecular	Volume
Carbon	14.8	H_2	14.3
Hydrogen	3.7	O_2	25.6
Chlorine	24.6	N_2	31.2
Bromine	27.0	Air	29.9
Iodine	37.0	CO	30.7
Sulfur	25.6	CO_2	34.0
Nitrogen	15.6	SO_2	44.8
Nitrogen in primary amines	10.5	NO	23.6
Nitrogen in secondary amines	12.0	N_2O	36.4
Oxygen	7.4	NH_3	25.8
Oxygen in methyl esters	9.1	H_2O	18.9
Oxygen in higher esters	11.0	H_2S	32.9
Oxygen in acids	12.0	COS	51.5
Oxygen in methyl ethers	9.0	CI_2	48.4
Oxygen in higher ethers	11.0	Br_2	53.2
Benzene ring: subtract	15	I_2	71.5
Naphthalene ring: subtract	30		

By permission, R. E. Treybal, *Mass-Transfer Operations*, p. 27, McGraw-Hill Book Co., Inc. (1955). Also see Ref. 63 for additional data.

20.34.1 Height Equivalent to a Theoretical Plate (HETP)

Distillation operations can best be expressed in terms of equilibrium relations and theoretical plates. Therefore, one of the correlating factors for various packings is the height of packing equivalent to a theoretical plate for the separation. Data for using this concept effectively are extremely meager and contain many uncertainties as far as general application is concerned. For this reason, the use of HETP is not popular. When good correlations are developed to predict HETP without test data, then this can be an effective means of expressing packing heights in distillation. Most HETP data have been obtained on small diameter (often laboratory size) columns using very small packings and operated at essentially total reflux or on moderate sized columns, but with limited systems. The scale-up of such data to industrial sizes is questionable.

When the designer does not have actual data from similar services (often broad classes of similarity may have to be assumed to arrive at what might be termed a reasonable and safe value for HETP), then it can be helpful to contact manufacturer's technical service departments for their recommendations (which they normally will develop from application files). Keep in mind that the HETP is also unique to the packing size and configuration; therefore, values obtained for one packing definitely do not accurately apply to another size or type. There is a rough relation based on recognizing that large size packing requires greater HETP than small size, but pressure drop is greater for the small packing.

There can be a significant difference between the conventional "particle" packing, as represented in most of the illustrations of Figure 20.6, and the HETP values for most of the structured packing in Figures 20.6V–20.OO and 20.6YFF. These latter types (structured packing) offer HETP values varying from about 5 to 14 in., which are lower (greater efficiency) than the random particles that have values varying from 12 in. to as much as 36 in. Plots of HETP values for various packings have been presented in previous charts and discussions in this chapter.

Due to the unique HETP characteristics, it is important to consult the manufacturer on the specific system and operating conditions involved. Care must be exercised in selecting these, or any other type of packing, since plugging with suspended solids, polymer formation on surfaces, and similar mechanical problem can influence performance and life of the packing system.

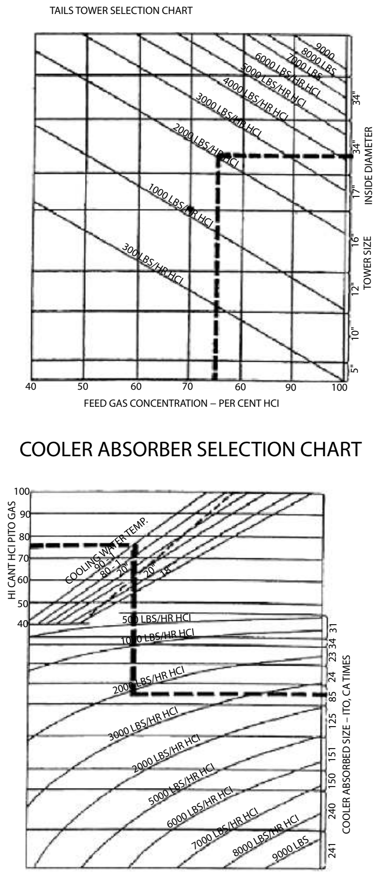
Many correlations for HETP have been limited to Raschig rings or Berl saddles [92], both being the least efficient for mass transfer and pressure drop when compared to the more sophisticated designs represented in Figure 20.6. The guidelines given in a later paragraph are adequate for most of these applications.

Cornell, Knapp, and Fair [87, 93, 94] proposed the use of the transfer-unit concept for distillation, where

$$H_{og} = H_g + m\left(\frac{L}{G}\right)_m H_l \tag{20.143}$$

where

H_{og} = height of overall gas-phase transfer unit, ft.
H_g = height of gas-phase transfer unit, ft.



To select correct size tails tower and cooler absorber, to meet your specific requirements, follow the dotted line in the two charts. For example, using a 75% concentration of feed gas, 90° cooling water, and absorbing 2,000 lbs./hr. HCl; a 92 tube cooler absorber would be required in conjunction with a 24" diameter tower. To produce acid over 20° Be to 22° Be , use 12-foot long tubes in all instances where cooling water temperature falls to left of black dotted line in cooler absorber selction chart. Use 9-foot long tubes where temperature point is to the right of the dotted line.

Figure 20.78 Preliminary selection charts for HCl towers. Used by permission of Falls Industries, Inc.

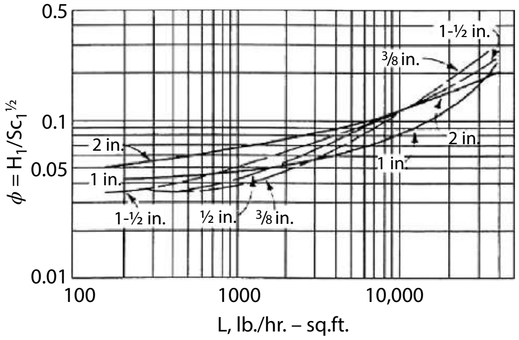


Figure 20.79 H_l correlation for various sizes of Raschig rings, Sc = Schmidt No. (N_{SC}). Reproduced by permission of the American Institute of Chemical Engineers, Cornell, *et al.*, Chemical Engineering Progress, V. 56, No. 8 © (1960) p. 68. All rights reserved.

H_l = height of liquid-phase transfer unit, ft.
m = slope of equilibrium curve
G = lb-mols gas/h/ft^2
L = lb-mols liquid/h/ft^2

$$H_l = \phi C_p \left(\frac{\mu_l}{\rho_l D_l}\right)^{0.5} \left(\frac{Z}{10}\right)^{0.15} \tag{20.144}$$

where

ϕ = correlation from Figures 20.78 and 20.79 for a given packing
C_p = correction factor for high rates, from Figure 20.80
μ_l = liquid viscosity, lb/h.ft
ρ_l = liquid density, lb/ft^3
D_l = liquid diffusion coefficient, ft^2/h
Z = height of packing, ft.

For Raschig rings:

$$H_g = \frac{\Psi S_{cg}^{0.5}}{(Lf_1 f_2 f_3)^{0.6}} \left(\frac{D}{12}\right)^{1.24} \left(\frac{Z}{10}\right)^{1/3} \tag{20.145}$$

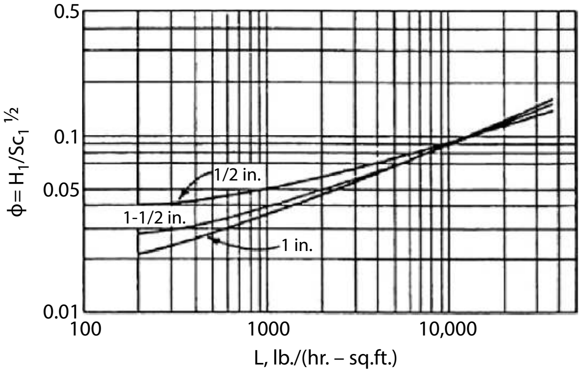


Figure 20.80 H_l correlation for various sizes of Berl saddles. Reproduced by permission of the American Institute of Chemical Engineers, Cornell, *et al.*, Chemical Engineering Progress, V. 56, No. 8 © p. 68. All rights reserved.

For Berl saddles:

$$H_g = \frac{\Psi S_{cg}^{0.5}}{(Lf_1f_2f_3)^{0.6}}\left(\frac{D}{12}\right)^{1.11}\left(\frac{Z}{10}\right)^{1/3} \tag{20.146}$$

where

Ψ = parameter for a given packing from Figures 20.81–20.83
S_{cg} = gas-phase Schmidt number = $\mu_g/\rho_g D_g$
μ_g = gas viscosity, lb/h. ft
ρ_g = gas density, lb/ft^3
D_g = gaseous diffusion coefficient, ft^2/h
D = column diameter, in.
Z = packed height, ft.
f_1 = $(\rho_1\ 2.42)^{0.16}$
f_2 = $(62.4/\ \rho_1)1.25$
f_3 = $(72.8/\sigma)^{0.8}$
μ_1 = liquid viscosity, lb/h. ft.
ρ_l = liquid density, lb/ft^3
σ = surface tension, dynes/cm

In Reference 92, the authors modified the equations for H_g and H_l as follows: (a) eliminate column diameter correction above 24 in. and (b) columns with good liquid distribution probably can allow elimination of the packing height correction.

Two separate investigators have evaluated the various correlation methods and reported that the Cornell *et al.* method is significantly better than others. These were 1- and 2-in. metal Pall rings and 1-in. ceramic Intalox saddles [26] and 3/4-, 1 ½-, and 3-in. Raschig rings [95].

Whitt [96] has correlated literature data of commercial size Raschig ring packing, as shown in Figure 20.84. The range of plotted data in indicated and the suggested design lines are good medians. The HTU values are for gas film controlling absorption systems, and the HETP data are for distillations at 760 mm Hg and below. These values should be usable for most pressure systems. The viscosity of the liquid ranges from 0.35 to 1.0 cP. The equation for the HETP line is [97]

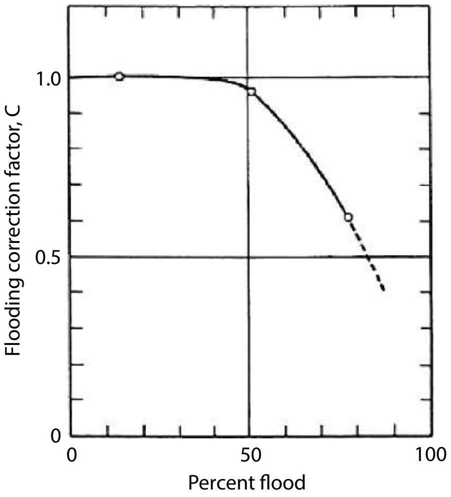


Figure 20.81 Liquid film correction factor for operation at high percent of flood. Reproduced by permission of the American Institute of Chemical Engineers, Cornell, *et al.*, Chemical Engineering Progress, V. 56, No. 8 © (1960) p. 68. All rights reserved.

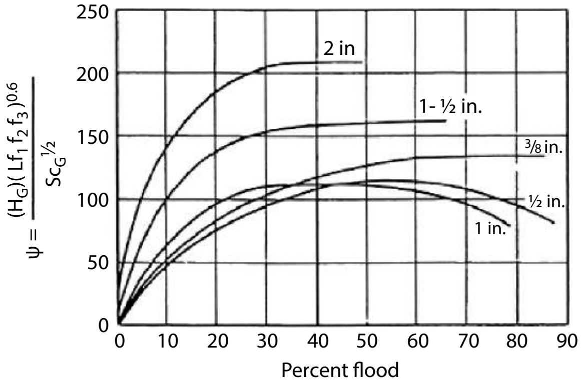


Figure 20.82 H_g, correlation for various sizes of Raschig rings. Reproduced by permission of the American Institute of Chemical Engineers, Cornell, *et al.*, Chemical Engineering Progress, V. 56, No. 8 © (1960) p. 68. All rights reserved.

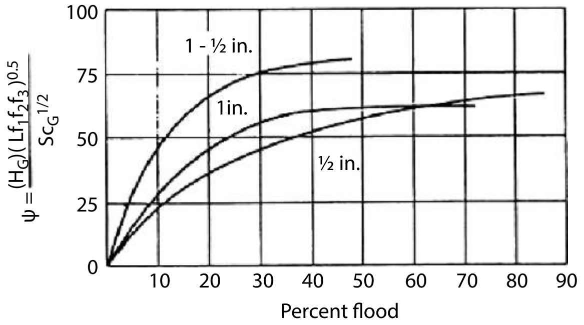
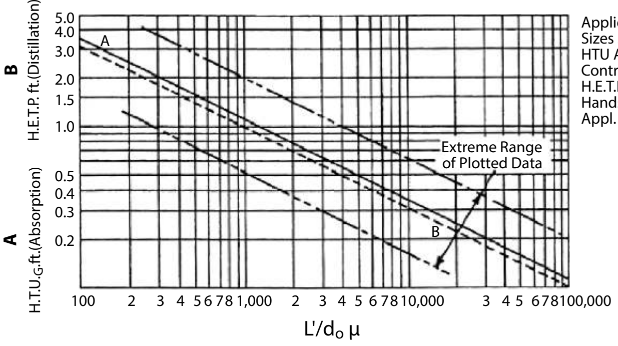


Figure 20.83 H_g, correlation for various sizes of Berl saddles. Reproduced by permission of the American Institute of Chemical Engineers, Cornell, *et al.*, Chemical Engineering Progress, V. 56, No. 8 © (1960) p. 68. All rights reserved.



Applies to Raschig Rings, Sizes 3/8" through 3". HTU Applies to Gas-Film Controlling Systems. H.E.T.P. Data froms C.H.G. Hands and F.R. Whitt, J. Appl. Chem. Mar.1, p.135(1951).

Figure 20.84 HETP and HTU correlation for tower random packings. Used by permission of Whitt, F. R., British Chemical Engineering, July (1955) p. 365. All rights reserved.

$$\text{HETP} = \frac{32}{\left(\dfrac{L'}{d_o\mu}\right)^{0.5}} \tag{20.147}$$

and for HTU_G

$$\text{HETP} = \frac{36}{\left(\dfrac{L'}{d_o\mu}\right)^{0.5}} \tag{20.148}$$

In general, for the same liquid and vapor rates, the HETP and HTU values for Berl saddles and others with a/ε^3 lower than Raschig rings should be lower. Correlating data is not available, except the tests of Teller [98], which indicate 1-in. Berl saddles have HETP values of 0.75 compared with 0.85 for 1-in. Raschig rings. Figure 20.83 shows Hg correlation for various sizes of Berl saddles.

The accuracy of establishing HETP values for new conditions in the same packing referenced to known values, as suggested by Planovski [99], has not been tested by other literature references.

$$(\text{HETP})_{\text{new}} = (\text{HETP})_{\text{known}} P^* G_{\text{new}} / P_{\text{new}} G^* \tag{20.149}$$

* Represents known conditions
P = system pressure, absolute

The operation of packed towers under vacuum conditions is not well defined in the literature. However, the work of Hands and Whitt [100] specifically evaluates several systems operating from 20 to 760 mm Hg abs. Their recommended limiting vapor velocity is

$$V_g = 0.065\left(\frac{\rho_L}{\rho_G}\right)^{0.5}, \text{ft/s} \tag{20.150}$$

or

$$L_m = 0.334(\text{Pressure, mm Hg.abs})^{0.5} \tag{20.151}$$

Operating values are recommended to be two-thirds of the limiting values. From the data, it appears that entrainment (internal may be more significant than external) becomes a limiting factor before the flooding values predicted by Figure 20.13C or D.

Eckert [101, 102] and others have examined data and presented considerable tabulations of HETP and HTU values for random packing in various organic-water, polar organic, and hydrocarbon systems. Some of these are now obsolete due to the older type/style packing tested; however, the results still have value as a base reference if it is recognized that the newer types/styles are more efficient, giving lower HETP and HTU values.

20.34.2 HETP Guide Lines

For industrial process equipment, some general guidelines in this undefined area are (for particle packing)

1. Never use HETP less than 12 in. if tower is 12 in. dia. or larger; for general assumption, use HETP = 1.5 to 2.0 ft.

2. Use HETP = H_{OG} or H_{OL} if other data are not available.
3. Use HETP = Column diameter (over 12 in. dia.) if no other information is available, up to 48 in. dia.

HETP values appear to vary somewhat within the distillation system, while certainly varying with the size and style of packing. In general, the larger, more open packing designs exhibit higher HETP values, while the smaller particle type packing and the compact styles exhibit significantly lower values for the same system. A brief guide to recently published values for various systems would be helpful in establishing the right order of magnitude for a system in design.

Kister [19] has tabulated published HETP data for a wide variety of process systems and proposes that using good experimental data to interpret for other systems can be just as effective (accurate) as using values calculated using more mass transfer techniques.

Influences on HETP Values

Various authors [19] and researchers have published details of factors that influence detail of HETP values. The lower the HETP value (i.e., number of feet (or inches) per theoretical plate/tray/stage), the more efficient is the packing (random or structured) for any particular separation system. Other influences include

1. Packing surface area/unit packing volume increases, resulting in more efficient packing performance (lower HETP). For structured packing, the narrower passageway between sheet components results in more efficient performance [19].
2. Uniformity of packing surface on a specific random element of packing [19].
3. Uniformity of liquid (vapor) distribution increases packing performance efficiency.
4. Liquid and vapor loadings have little effect on HETP for random packing up, to the point between loading and flooding.
5. For structured packing, as loadings increase, HETP increases. This is more pronounced for wire-mesh types, and the effect is less for the corrugated sheet packing [19].
6. Pressure above 1 or 2 psia has small effects on HETP for random and structured packings. For high vacuum, the data are not totally firm as to the consistency of effects; generally, HETP increases (efficiency becomes lower).

20.34.3 Transfer Unit

The transfer unit concept is also applicable to distillation in packed towers. Height of the packing required is

$$Z = N_{OG}\, H_{OG} \qquad (20.152)$$

For usual applications in the rectifying section where the number of transfer units N_{OG} is given by

$$N_{OG} = \int_{y_1}^{y_2} \frac{dy}{y^* - y} \qquad (20.153)$$

For the stripping section, usually

$$Z = N_{OL}\, H_{OL} \qquad (20.154)$$

The height of the transfer unit has not been satisfactorily correlated for a wide variety of systems. If pilot plant or other acceptable data are available to represent the system, then the height of packing can safely be scaled up to commercial units. If such data are not available, rough approximations may be made by determining H_G and H_L as for absorption, and combining to obtain an H_{OG} (Ref. 88). This is only very approximate. In fact, it is due to lack of any volume of data obtained from commercial units that many potential applications of packed towers are designed as trays towers.

$$N_{OL} = \int_{x_1}^{x_2} \frac{dx}{x^* - x} \text{ for liquid concentration gra dients} \tag{20.155}$$

Based on the two-resistance theory [103]

$$(HTU)_{OG} = (HTU)_G + (HTU)_L \tag{20.156}$$

If, over a single stage, the equilibrium curve and the operating lines are assumed straight to be, then [103]

$$HETP = (HTU)_{OG}(\ln\lambda)/(\lambda - 1); \text{ for } \lambda \neq 1.0 \tag{20.157}$$

$$HETP = (HTU)_{OG}; \text{ for } \lambda = 1.0 \tag{20.158}$$

L_M/G_M is the internal reflux ratio that is the slope of the operating line, in either the rectifying or stripping sections. The value of L_M/G_M usually will not vary significantly from 1.0. $\lambda = mG_M/L_M$ is the ratio of the two slopes and remains close to 1.0 for different separation systems [103]. For these conditions, HETP ≅ $(HTU)_{OG}$ and HETP becomes an important design variable. For $\alpha \gg 1.0$ for easy separations and λ is less than or much greater than 1.0. See Figures 20.85 and 20.86, which illustrate the effect on HETP of under-irrigating the bed by the distributor.

A minimum HETP or HTU represents a maximum separation efficiency, with α representing the relative volatility, i.e., vapor and liquid phase compositions of the more volatile component in a binary system:

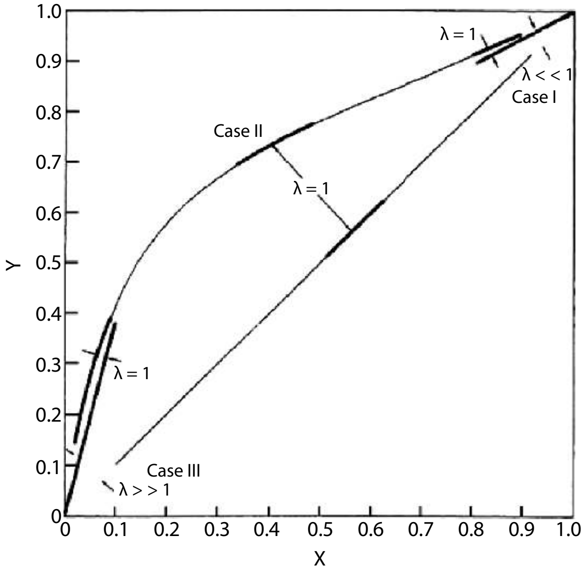


Figure 20.85 Vapor-liquid equilibrium showing λ and application cases referred to in the text. Used by permission, Koshy, T. D. and Rukovena, F., Jr., Hydrocarbon Processing V. 65, No. 5 (1986) p. 64. All rights reserved.

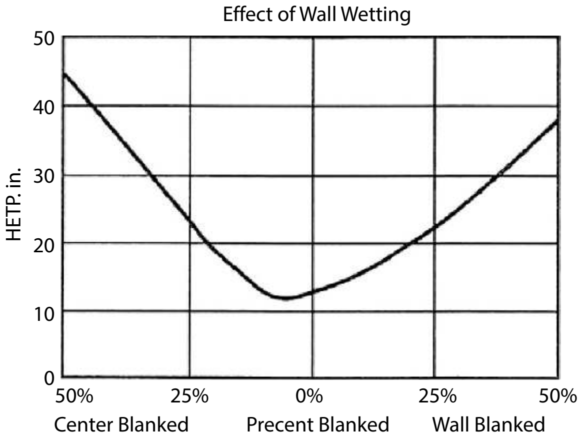


Figure 20.86 Effect of under-irrigating the wall area of a column using random packing. Note: % blanked refers to packing cross-sectional area not irrigated. At 0% blanked, the best HETP is found. To the right of 0 towards the wall (50% blanked), the HETP becomes poorer until the wall is reached. For the center section (to the left) of the tower from the 0% to 50% of the center area of tower area blanked, the poorer HETP is again found. Reproduced by permission of The American Institute of Chemical Engineers, Chemical Engineering Progress, Bonilla, J. A., V. 89, No. 3 (1993), p. 47; and by special permission of Fractionation Research, Inc. All rights reserved.

$$y = \alpha x[1+(\alpha - 1)x] \tag{20.159}$$

The following examples are presented here by permissions of the authors [103] and Hydrocarbon Processing:

Case I: γ is much less than 1. The circumstances when this happens are [103]:

- High purity of the more volatile component, i.e., $m \longrightarrow 1/\alpha$
- High reflux ratio, i.e., $L_M/G_M \longrightarrow 1.0$
- Both foregoing conditions mean $\lambda \longrightarrow 1/\alpha$

Case II. $\lambda = 1$. The following situations will produce this condition [103]:

- Very low reflux ratio for high purity rectification, i.e., $x \longrightarrow 1.0$
- Very high reflux ratio for high purity stripping, i.e., $x \longrightarrow 0$.
- Total reflux for a symmetric separation. Note that the term "symmetric separation" is used here to mean that on a McCabe–Thiele diagram, the liquid phase compositions of the overhead product and bottom product are roughly equidistant from 0.5.

Case III. λ is much greater than 1. The circumstances when this occurs are [103]:

- High purity of bottoms products, i.e., $x \rightarrow 0$ and $m \rightarrow \alpha$
- Low L/V approaching total reflux, i.e., $L/V \rightarrow 1.0$
- Both foregoing conditions mean that $\lambda \rightarrow \alpha$

where

G_M = gas flow rate, kg-mols/h or lb mols/h.
HETP = height equivalent to a theoretical plate, m or ft.
HTU = height of a transfer unit, m or ft.
L_M = liquid rate, kg-mol/h or lb mols/h
m = slope of the equilibrium line
x = mol fraction of the more volatile component in the liquid phase
y = mol fraction of the more volatile component in the vapor phase

Greek Symbols

α = relative volatility between components 1 and 2
$\lambda = mG_M/L_M$ = ratio of the slopes of equilibrium and operating lines

Subscripts

G = gas phase
L = liquid phase
OG = overall gas phase

Packed column performance can use either the HETP or HTU concepts; the HTU is somewhat more complicated but no more correct than the HETP concept. The latter adapts itself to direct use from tray-by-tray digital computer calculations and is thereby a little more direct.

The packed column has been quite useful in distillation, stripping, and absorption processes and has become competitive with many types of distillation tray designs or types/styles.

Bolles and Fair [104] present an analysis of considerable data in developing a mass-transfer model for packed tower design; however, there is too much detail to present here.

Example 20.11: Transfer Units in Distillation

A benzene–toluene mixture is to be separated in a tower packed with 1-in. Berl saddles. The feed is 55.2 mol% (liquid feed saturated), and an overhead of 90 mol% benzene and bottoms of not more than 24 mo% benzene are desired. Using the data of Ref. [105], plotted in Figure 20.87, determine the number of transfer units in the rectifying and stripping sections using a reflux ratio (reflux to product, L/D) of 1.35.

Solution

Referring to Figure 20.87 for the graphical solution:

Rectifying section operating line slope

$$\frac{R}{R+1} = \frac{1.35}{1.35+1} = 0.576$$

Note that point 7 can be determined by the intersection of the rectifying operating line and the feed condition line 8-7.

Establish the location of the feed, bottoms, and overhead compositions on the graph. Draw in the operating lines as for a distillation in a tray column.

To establish the transfer units, draw in line A–B–C so that it is always halfway vertically between the equilibrium line and the operating line, making dimension 1–2 equals to 2–3. Begin drawing the transfer units at the overhead product 4, such that 4–9 equals 9–5, then drop vertically to the operating line and repeat the process, always making the line A–B–C bisect the horizontal portion of the step. At the feed point, restart the stepwise process if the transfer unit step does not terminate at the feed point 7.

For this example, the number of transfer units is

Rectifying	2.9 ±
Stripping	2.6 ±
Total	5.5±

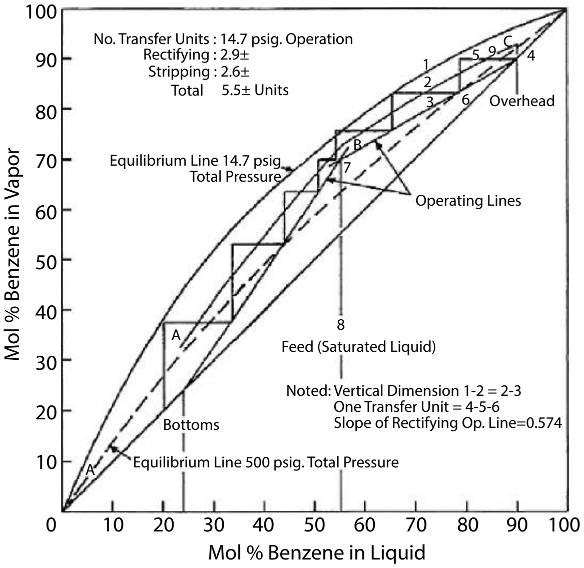


Figure 20.87 Vapor–liquid equilibrium (data only), benzene–toluene. Reproduced by permission of the American Institute of Chemical Engineers, Griswold, J., Anders, D., and Klein, V. A., Trans. A.I.Ch.E, V. 39© (1943) p. 223. All rights reserved.

The reboiler for the column is in addition to this; however, the bottoms were specified as being the inlet from the reboiler. For most purposes, the reboiler can be considered to be one additional transfer unit.

Alternative

An alternative method to determining the number of transfer units is the graphical integration of $dy/(y^* - y)$. The procedure is basically the same as for absorbers, that is,

1. For assumed values of x (mol fraction of component under consideration in liquid) from bottoms to overhead, read values of y (vapor under operating conditions corresponding to x) and values of y^* (vapor in equilibrium with x) from the equilibrium line.
2. Calculate $1/(y^* - y)$ at each selected point, thus:

x	y*	y	y* - y	1/(y* - y)
0.24	0.44	0.24	0.20	5.0
0.30	0.52	0.328	0.192	5.21

3. Plot y from y bottoms to y overhead versus $1/(y^* - y)$. The position of y-feed can be noted on the graph, and the integration so arranged as to reveal the split between rectifying and stripping transfer units. The total number by this method should check closely with the graphical stepwise method.

Height of Transfer Unit

The height of a transfer unit for this system is not available; therefore, it may be roughly approximated by the method of additive H_G and H_L, which is questionable at best, or the approximation of 2 ft. for HTU may be used. The latter is just about as reliable as the former.

Then: height of tower packing, using 1-in. Berl saddles:

From transfer unit = (2) (5.5) =	11 ft
Allowances: for top distribution:	2 ft
Subtotal	13 ft
Extra, 20%	2.5 ft
Total Z	15.5 ft , use 16 ft.

The tower must be designed for throughput-diameter determined and pressure drop established.

Check Theoretical Plate Basis

To determine HETP by approximate method, see Table 20.47 for benzene–toluene:

HETP = 1.0 to 1.5 ft
Select HETP = 1.25 ft
Safety factor suggested = 2 in., in any case a value not less than 1.25

Therefore, use
HETP = 24 in. = 2 ft.

From Figure 20.87, it is evident that the number of theoretical plates and number of transfer units are not the same. When stepped off, the number of theoretical plates is 6+.

Height of packing = (6) (2) =	12 ft
Allowance for distribution =	2 ft
Total	14 ft

Use: 14-ft packing, 1-in. Berl saddles.

Because the 16 ft of packing by the HTU method is larger, this would be the recommended safe height to use. For comparison, note the relative increase in the number of transfer units if the operation were at a higher pressure as shown by the dotted line for 500 psig.

Strigle [22] discusses packed column efficiency (HETP) in considerable detail. Most of his published data refer to the work of Norton Chemical Process Products Corp.

A Norton [22] correlation for modern, random dumped packings used for distillation up to 200 psia is (use high-performance internal distributors and supports) from surface tension of 4 dynes/cm but less than 33, and liquid viscosity of at least 0.83 to 0.08 cP but not greater:

$$\ln \text{HETP} = n - 0.187 \ln \sigma + 0.213 \ln(\mu) \tag{20.160}$$

Values of n in the equation for selected specific packings are given in Table 20.48 for random packing and Table 20.49 for structured packing.

Table 20.47 HETP estimates for distillation applications.

Contact manufacturers for specific design recommendations			
System	**Pressure Range If Known**	**General Packing Type/Style/Make**	**Estimating HETP* Ft. or In. Marked**
Iso-octane/toluene	100 mm Hg	Hy-Pak No. 2	2.0–2.7
Same	740 mm Hg	Hy-Pak No. 1	0. 7–2.7
Para/ortho xylene	740 mm Hg	Metal Intalox No. 25	0.8–1.3
Same	740 mm Hg	Metal Intalox No. 40	1.3–1.55
Same	740 mm Hg	Metal Intalox No. 50	1.75–2.15
Chlorinated HC	Vacuum	Metal Intalox No. 25	2
Chlorinated HC	Vacuum	Metal Intalox No. 40	2.4
Chlorinated HC	Vacuum	Metal Intalox No. 50	3.5
Iso-octane/toluene	740 mm Hg	Pall Ring, 1 in. Metal	1.0–2.0
Iso-octane/toluene	740 mm Hg	Pall Ring, 1½ in. Metal	0.75–1.0 (3.5)**
Iso-octane/toluene	740 mm Hg	Pall Ring, 2 in. Metal	1.5–2.2
Methanol/water	740 mm Hg	Pall Ring, 1 in. Metal	0.65–0.8 (1.2)**
Isopropanol/water	740 mm Hg	Pall Ring, 1 in. Metal	0.6–1.5
Benzene/toluene	740 mm Hg	Pall Ring, 1 in. Metal	1.0–1.5
Acetone/water	740 mm Hg	Pall Ring, 1 in. Metal	0.9–1.2 (1.4)**
Same	–	Flexirings, 1 in. Metal	1.6–1.8 (2.3)**
Same	–	Flexirings, 2 in. Metal	1.8–2.2 (2.4)**
Same	–	Koch Sulzer Metal	0.45–0.9
Light hydrocarbon	400 psia	Goodloe Metal	approx. 0.75
Propane/butane	235 psia	Goodloe Metal	approx. 0.80
Chlorobenzene/ethylbenzene	50 mm Hg	Montz structured metal	5 in.–17 in.
Chlorohexane/n-heptane	1 atm	#2 Nutter Ring	22 in.–30 in.
Chlorohexane/ n-heptane	5 psia	#2 Nutter Ring	22 in.–30 in.
Various sys.	vacuum	Goodloe metal, various	5 in.–8 in.
Iso-Octane/toluene	1 atm	Cascade Mini-ring, #3	22 in.–28 in.
Iso-Octane/toluene	1 atm	Cascade Mini-ring, #2	18 in.–24 in.
h-naphtha/light gas oil	unknown	Gempak, ½ in. crimp	13 in.–15 in.
h-naphtha/light gas oil	unknown	Gempak, 1 in. crimp	22 in.–27 in.
h-naphtha/light gas oil	unknown	Gempak, ¼ in. crimp	8 in.–10 in.
Ethylene dichloride/benzene	1 atm	ACS-X Mesh	4 in.–9 in.

(*Continued*)

Table 20.47 HETP estimates for distillation applications. (*Continued*)

Contact manufacturers for specific design recommendations			
System	**Pressure Range If Known**	**General Packing Type/Style/Make**	**Estimating HETP* Ft. or In. Marked**
Methylcyclohexane/toluene	1 atm	ACS-X-200 Mesh	3.5 in.–12 in.
Unknown	unknown	Koch structured Flexipac	17 in.
General/average	unknown	Koch/Sulzer (R) structured	3 in.–9 in.
Ortho/para-	16 mm Hg abs	Koch/Sulzer (R) structured	5 in.–16 in.
Ortho/para-	100 mm Hg abs	Koch/Sulzer (R) structured	4.5 in.–8 in.

*Based on industrial data or commercial sized tests, note some values in inches.
**At very low gas rates.
Data for table compiled from respective manufacturer's published literature.

Table 20.48 Constant n for HETP correlation*, random packing.

Tower packing	**Value of n**
#25 IMTP® packing	1.13080
#40 IMTP® packing	1.37030
#50 IMTP® packing	1.56860
1 in. Pall Ring	1.13080
1 ½ in Pall Ring	1.39510
2 in. Pall Ring	1.65840
1 in. Intalox® Saddle	1.13080
1 ½ in. Intalox® Saddle	1.41570
2 in. Intalox® Saddle	1.72330

*Use with Equation 14-125.
**IMTP and Intalox are registered names of Norton Chemical Process Products Corp.
Used by permission from Strigle, R. E, Jr., *Packed Tower Design and Applications: Random and Structured Packings;* 2nd Ed. Gulf Pub. Co. (1994).

Table 20.49 Constant n for HETP correlation*, for Intalox structured packing.

Packing size	**Value of n**
1T	0.76830
2T	1.01280
3T	1.38680

*Use with Equation 14-125.
**IMTP and Intalox are registered names of Norton Chemical Process Products Corp.
Used by permission from Strigle, R. E, Jr., *Packed Tower Design and Applications: Random and Structured Packings;* 2nd Ed. Gulf Pub. Co. (1994).

Strigle presents typical separation efficiency ranges for Intalox® metal tower packing, for systems with relative volatility not greater than 2.0.

Packing size	HETP (ft)
# 25	1.2 to 1.6
#40	1.5 to 2.0
#50	1.8 to 2.4

where

σ = surface tension, dynes/cm
μ = liquid viscosity, cP
n = constant for HETP Equation 20.152

Data for table compiled from respective manufacturer's published literature.

The summary of HETP values of Vital [106] for various types and sizes of packings is believed to be referenced to typical industrial distributors for the liquid. This variation can influence the value of HETP in any tabulation; the effect of distributor design is discussed in an earlier section of this chapter. Porter and Jenkins [107] developed a model to improve the earlier models of Bolles and Fair from about 25% deviation to about a 95% confidence using a 20% factor of safety [22].

Strigle [22] recommends:

1. For easy separations (less than 10 theoretical stages), a 20% design safety factor can be applied to a typical HETP value.
2. For separations of 15 to 25 theoretical stages, a 16% design safety factor should be applied to the HETP.
3. For very difficult separations, the design HETP should be carefully evaluated by calculation and actual data when available.
4. HETP values for random dumped packing have been found to be 25% greater at a greater viscosity than a lower viscosity, i.e., viscosity change from 0.15 to 0.44 cP.

Case Study-1

Absorption of Hydrocarbons with Lean Oil Using UniSim Design® R460.2

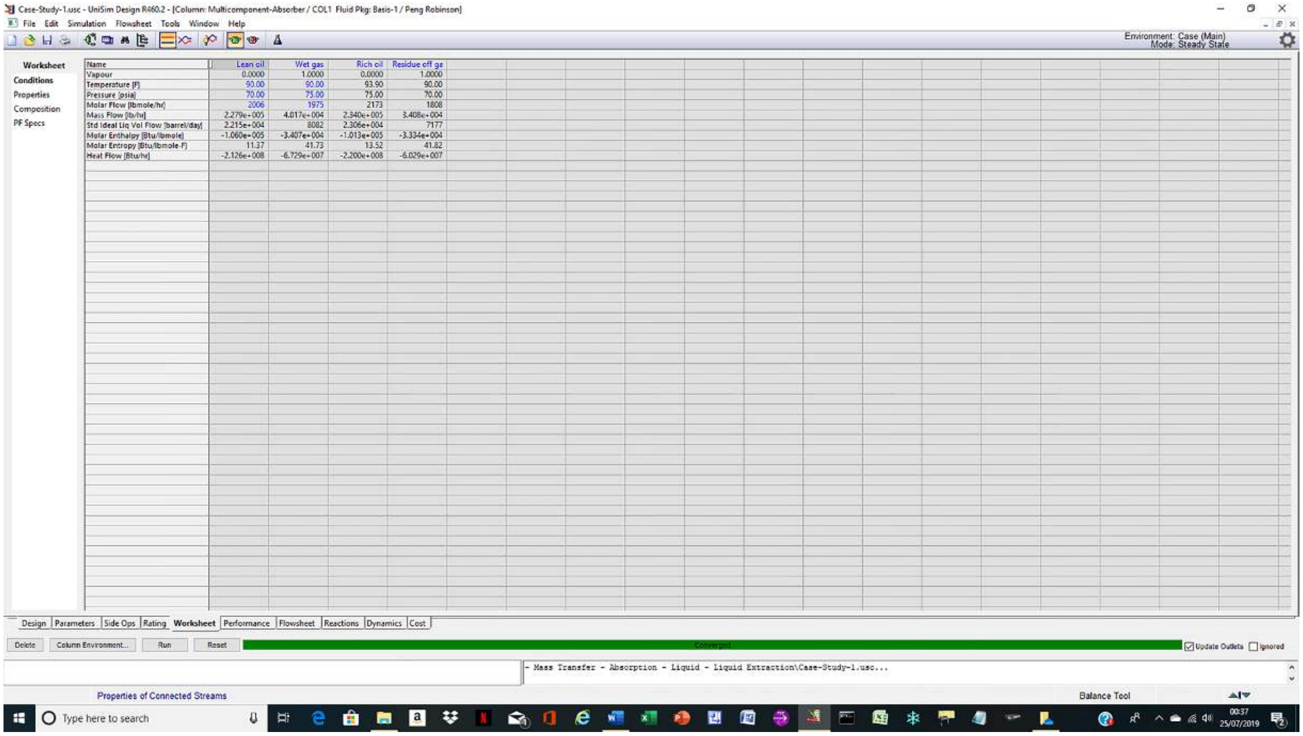
The gas stream shown in Table 20.50 is fed to an isothermal thermal absorber operating at 90°F and 75 psia; 90% of the n-butane is to be removed by contact with a lean oil stream consisting of 98.7 mol % non-volatile oil and light components shown in Column 2 of Table 20.50. Using UniSim Design to determine the product streams and the required number of theoretical stages

Solution

UniSim Design® R460.2 (Case-study -20.1.usc) was used to simulate the multicomponent absorption of wet gas using lean oil and Figures 20.88 to 20.92 show the snap shots of the parameters. Figure 20.93 shows the process flow diagram of the multicomponent absorption unit, and Tables 20.51 to 20.56 show the results of the simulation.

Table 20.50 Calculations summaries for Case Study-1.

					Off-Gas					
Component	**90°F K 75 psia**	**Feed Gas In (mols/hr)**	**Lean oil in (mol. Fraction)**	**Initial Estimate Of Net Amount Absorbed (mols/hr)**	**Initial Estimate (mols/hr)**	**After First Iteration (mols/hr)**	**After Second Iteration (mols/hr)**	**After Third Iteration (mols/hr)**	**Lean Oil In (mols/hr)**	**Rich Oil Out (mols/hr)**
Methane	42.5	1,639.2	–	22.6	1,616.6	1,597.5	1,598.4	1,598.4	–	40.8
Ethane	7.3	165.8	–	13.3	152.5	141.2	141.8	141.8	–	24.0
Propane	2.25	94.9	–	24.67	70.23	49.84	50.76	50.76	–	44.14
i-Butane	0.88	17.8	0.001	11.83	5.97	3.10	3.13	3.13	1.91	16.58
n-Butane	0.65	33.6	0.002	30.24	3.36	3.36	3.36	3.36	3.83	34.07
i-Pentane	0.28	7.9	0.004	7.9	0	2.08	2.03	2.03	7.66	13.53
n-Pentane	0.225	15.8	0.006	15.8	0	2.51	2.44	2.44	11.49	24.85
Heavy Oil	0	–	0.987	0	0	0	0	0	1,889.83	1,889.83
		1,975.0	1.000	126.34	1,848.66	1,799.57	1,801.92	1,801.92	1,914.72	2,087.78

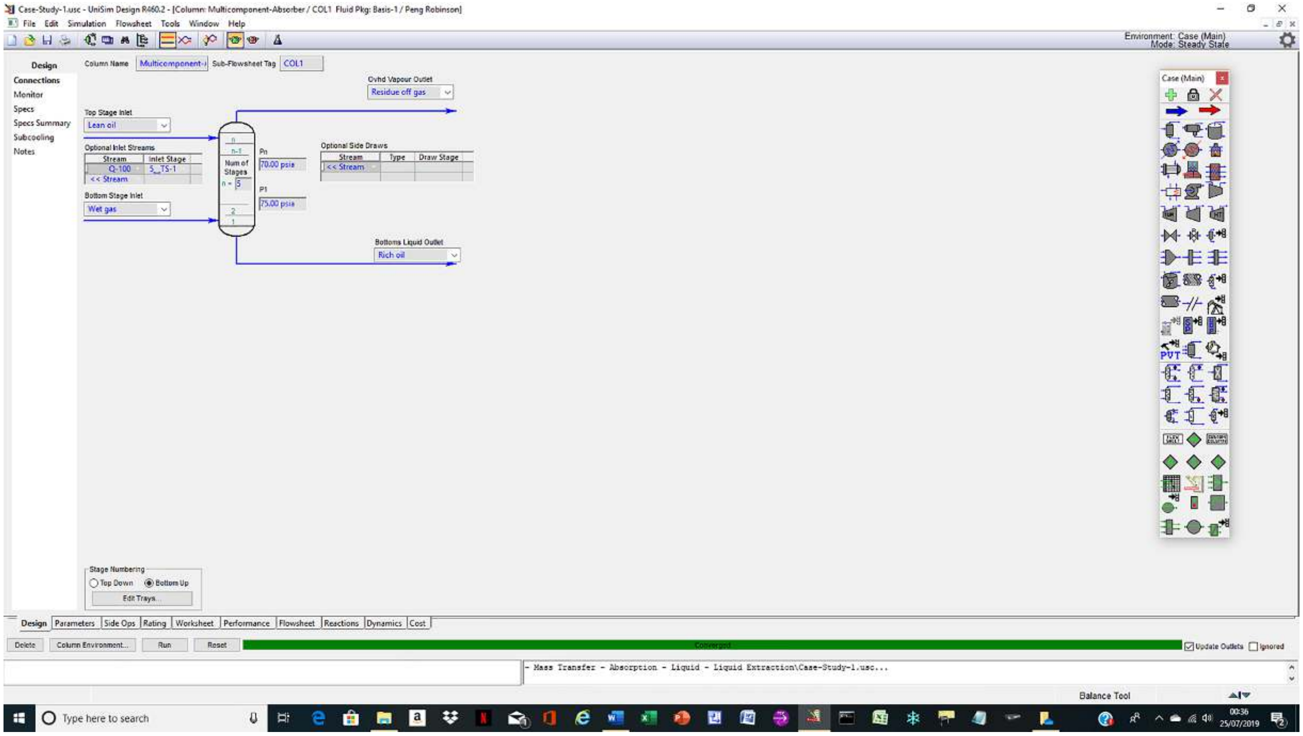


Case-Study-1.usc - UniSim Design R460.2 - [Column: Multicomponent-Absorber / COL1 Fluid Pkg: Basis-1 / Peng Robinson]

Name	Lean oil	Wet gas	Rich oil	Residue off ga
Vapour	0.0000	1.0000	0.0000	1.0000
Temperature [F]	90.00	90.00	93.90	90.00
Pressure [psia]	70.00	75.00	75.00	70.00
Molar Flow [lbmole/hr]	2006	1975	2173	1808
Mass Flow [lb/hr]	2.279e+005	4.017e+004	2.340e+005	3.408e+004
Std Ideal Liq Vol Flow [barrel/day]	2.215e+004	8082	2.306e+004	7177
Molar Enthalpy [Btu/lbmole]	-1.060e+005	-3.407e+004	-1.013e+005	-3.334e+004
Molar Entropy [Btu/lbmole-F]	11.37	41.73	13.52	41.82
Heat Flow [Btu/hr]	-2.126e+008	-6.729e+007	-2.200e+008	-6.029e+007

```
- Mass Transfer - Absorption - Liquid - Liquid Extraction\Case-Study-1.usc...
```

Figure 20.88 A snapshot of the worksheet window showing the streams of Lean oil, Wet gas, Rich oil, and Residue Off gas (courtesy of Honeywell Process Solution. UniSim Design® R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



Case-Study-1.usc - UniSim Design R460.2 - [Column: Multicomponent-Absorber / COL1 Fluid Pkg: Basis-1 / Peng Robinson]

Column Name: Multicomponent-A
Sub-Flowsheet Tag: COL1
Top Stage Inlet: Lean oil
Bottom Stage Inlet: Wet gas
Ovhd Vapour Outlet: Residue off gas
Bottoms Liquid Outlet: Rich oil
Num of Stages: n = 5
Pn: 70.00 psia
P1: 75.00 psia

Optional Inlet Streams

Stream	Inlet Stage
Q-100	5__TS-1
<< Stream	

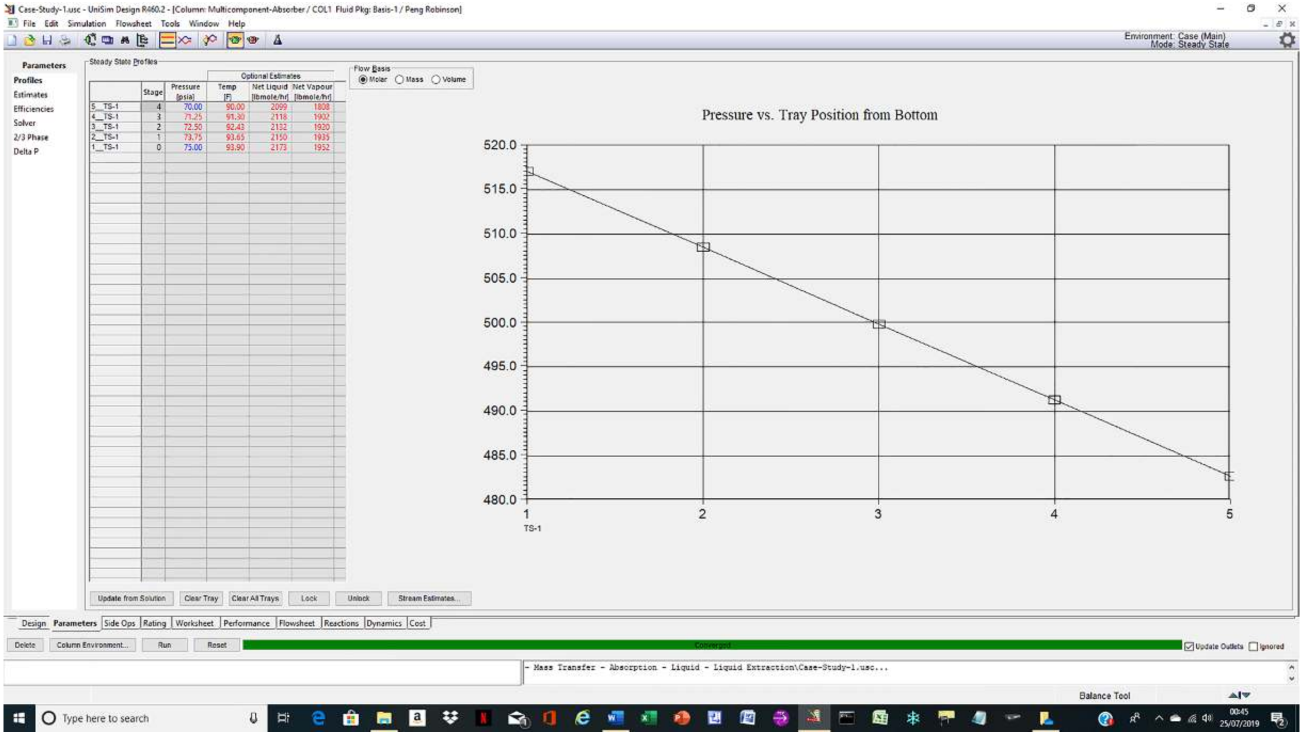
Optional Side Draws

Stream	Type	Draw Stage
<< Stream		

Stage Numbering: ○ Top Down ● Bottom Up

```
- Mass Transfer - Absorption - Liquid - Liquid Extraction\Case-Study-1.usc...
```

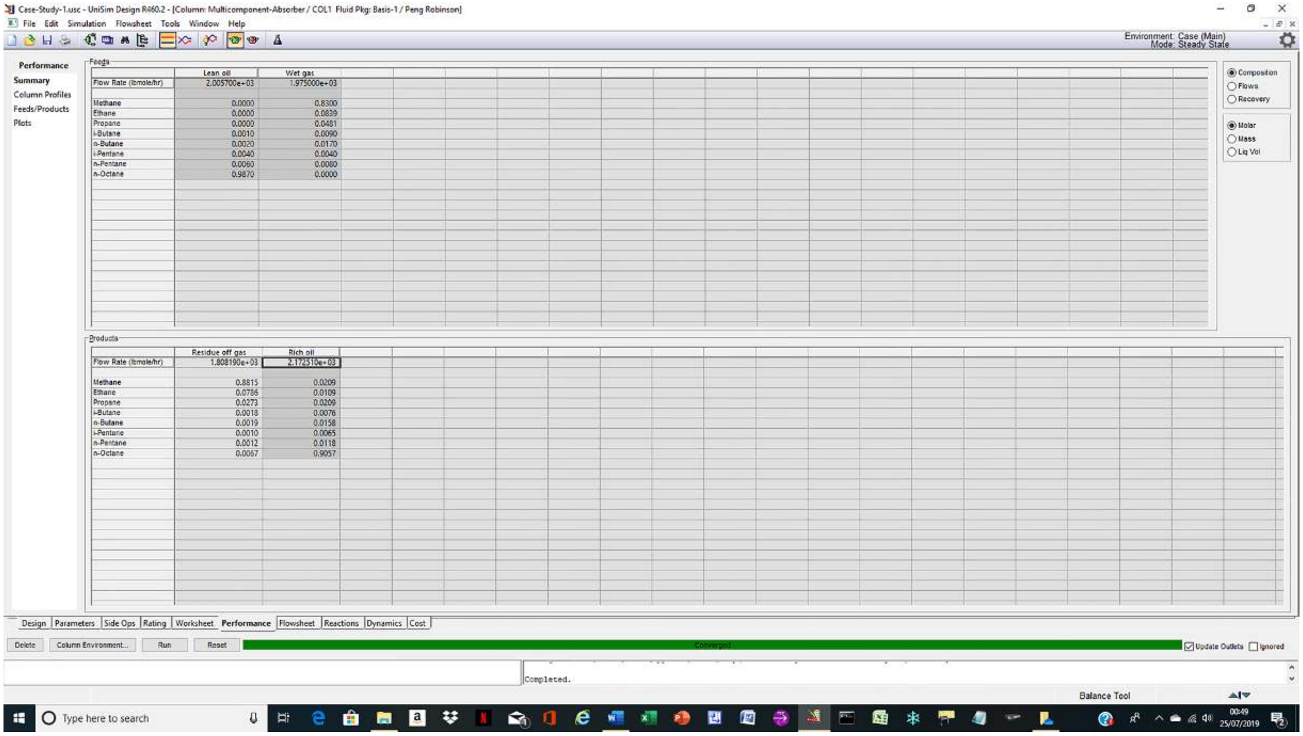
Figure 20.89 A snapshot of the Design window in the Connection page (courtesy of Honeywell Process Solution. UniSim Design® R460.2, Honeywell®, and UniSim®, are registered trademarks of Honeywell International Inc.).



	Stage	Pressure [psia]	Optional Estimates Temp [F]	Net Liquid [lbmole/hr]	Net Vapour [lbmole/hr]
5__TS-1	4	70.00	90.00	2099	1808
4__TS-1	3	71.25	91.30	2118	1902
3__TS-1	2	72.50	92.43	2132	1920
2__TS-1	1	73.75	93.65	2150	1935
1__TS-1	0	75.00	93.90	2173	1952

- Mass Transfer - Absorption - Liquid - Liquid Extraction\Case-Study-1.usc...

Figure 20.90 A snap-shot of the pressure vs. Tray position in the Parameter page (courtesy of Honeywell Process Solution. UniSim Design® R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



Feeds

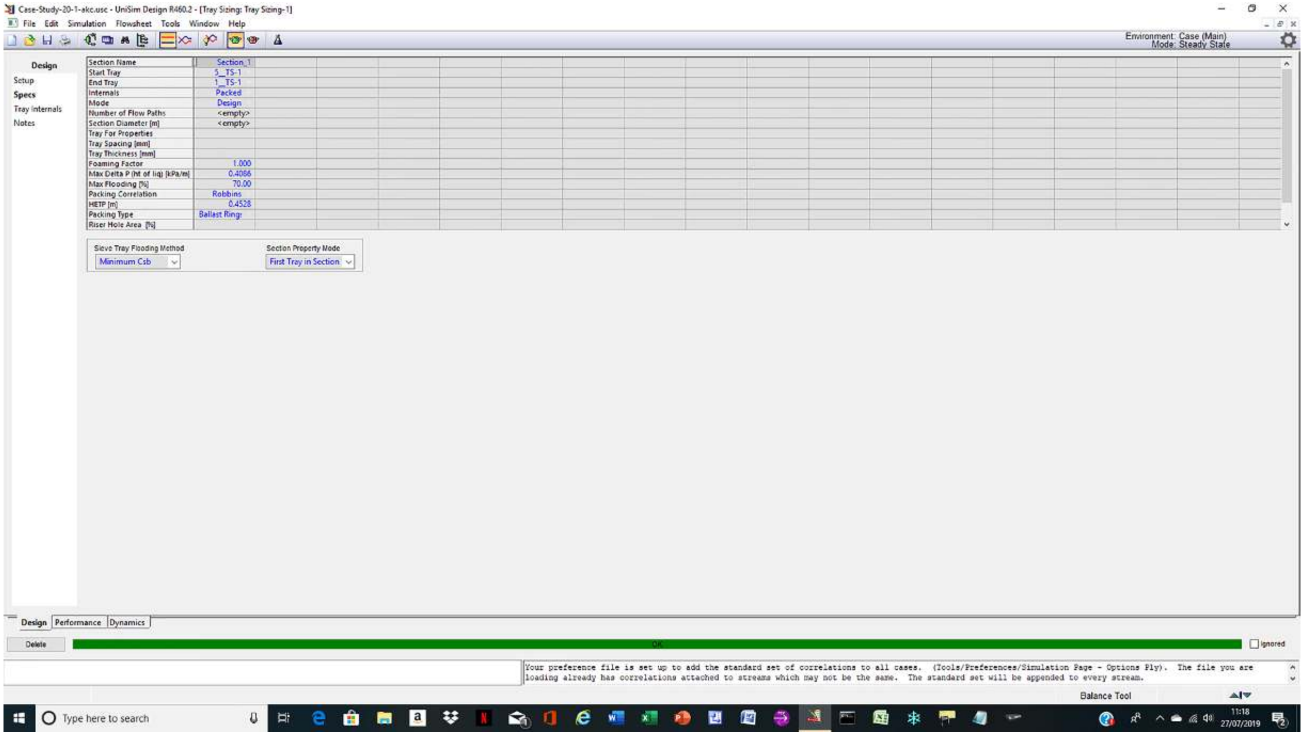
	Lean oil	Wet gas
Flow Rate (lbmole/hr)	2.005700e+03	1.975000e+03
Methane	0.0000	0.8300
Ethane	0.0000	0.0839
Propane	0.0000	0.0481
i-Butane	0.0010	0.0090
n-Butane	0.0020	0.0170
i-Pentane	0.0040	0.0040
n-Pentane	0.0060	0.0080
n-Octane	0.9870	0.0000

Products

	Residue off gas	Rich oil
Flow Rate (lbmole/hr)	1.808190e+03	2.172510e+03
Methane	0.8815	0.0209
Ethane	0.0786	0.0109
Propane	0.0273	0.0209
i-Butane	0.0018	0.0076
n-Butane	0.0019	0.0158
i-Pentane	0.0010	0.0065
n-Pentane	0.0012	0.0118
n-Octane	0.0057	0.9057

Completed.

Figure 20.91 A snapshot of the performance window in the Summary page (courtesy of Honeywell Process Solution, UniSim Design® R460.2, Honeywell®, and UniSim ® are registered trademarks of Honeywell International Inc.).



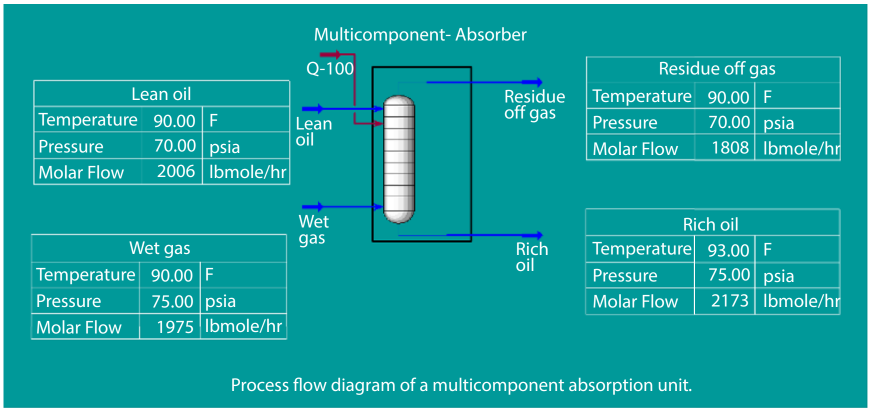
Section Name	Section_1
Start Tray	5__TS-1
End Tray	1__TS-1
Internals	Packed
Mode	Design
Number of Flow Paths	<empty>
Section Diameter [m]	<empty>
Tray For Properties	
Tray Spacing [mm]	
Tray Thickness [mm]	
Foaming Factor	1.000
Max Delta P (ht of liq) [kPa/m]	0.4086
Max Flooding [%]	70.00
Packing Correlation	Robbins
HETP [m]	0.4528
Packing Type	Ballast Ring:
Riser Hole Area [%]	

Sieve Tray Flooding Method: Minimum Csb

Section Property Mode: First Tray in Section

```
Your preference file is set up to add the standard set of correlations to all cases.  (Tools/Preferences/Simulation Page - Options Ply).  The file you are
loading already has correlations attached to streams which may not be the same.  The standard set will be appended to every stream.
```

Figure 20.92 A snapshot of the Design window in the Specs page showing HETP and Packing Type (courtesy of Honeywell Process Solution, UniSim Design® R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



Lean oil		
Temperature	90.00	F
Pressure	70.00	psia
Molar Flow	2006	lbmole/hr

Wet gas		
Temperature	90.00	F
Pressure	75.00	psia
Molar Flow	1975	lbmole/hr

Residue off gas		
Temperature	90.00	F
Pressure	70.00	psia
Molar Flow	1808	lbmole/hr

Rich oil		
Temperature	93.00	F
Pressure	75.00	psia
Molar Flow	2173	lbmole/hr

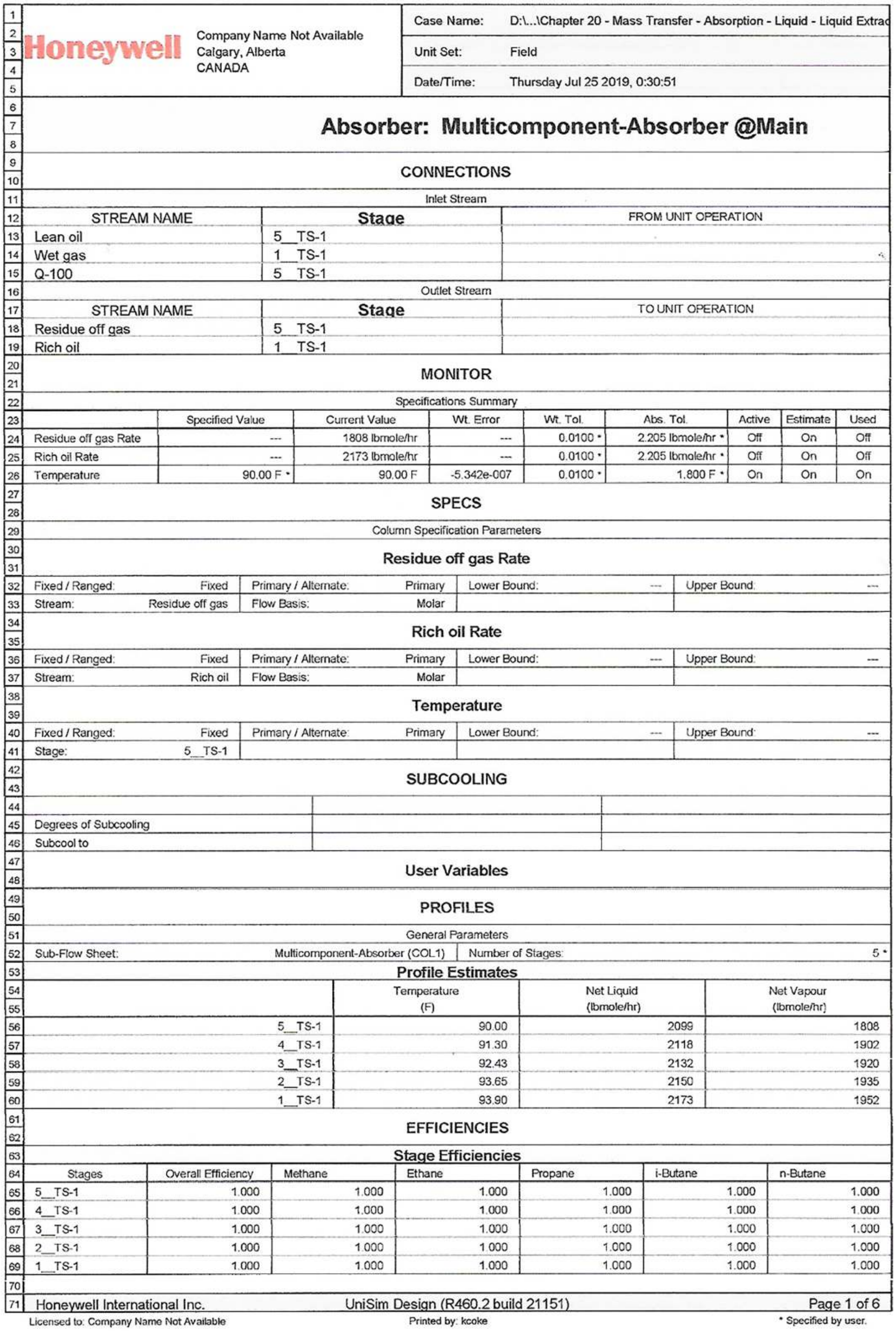
Process flow diagram of a multicomponent absorption unit.

Figure 20.93 A snapshot of the flow diagram of a multicomponent absorption unit (courtesy of Honeywell Process Solution, UniSim Design® R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).

20.35 Liquid–Liquid Extraction

Liquid–liquid extraction is a separation technique used for a wide range of applications in the refinery and chemical process industries. Where distillation is based on boiling point differences, extraction separates components based on their relative solubilities in two immiscible liquids. This technique is employed for separation applications that would not be cost-effective, or even possible, with distillation. It depends upon contacting the original mixture or solution in the liquid state, with a second liquid (the solvent) that is either immiscible or only partially miscible with the original solution. This is chosen such that the desired component of the mixture preferentially transfers or is

Table 20.51 Simulation results of multicomponent absorber.



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	D:\...\Chapter 20 - Mass Transfer - Absorption - Liquid - Liquid Extrac
Unit Set:	Field
Date/Time:	Thursday Jul 25 2019, 0:30:51

Absorber: Multicomponent-Absorber @Main

CONNECTIONS

Inlet Stream

STREAM NAME	Stage	FROM UNIT OPERATION
Lean oil	5__TS-1	
Wet gas	1__TS-1	
Q-100	5__TS-1	

Outlet Stream

STREAM NAME	Stage	TO UNIT OPERATION
Residue off gas	5__TS-1	
Rich oil	1__TS-1	

MONITOR

Specifications Summary

	Specified Value	Current Value	Wt. Error	Wt. Tol.	Abs. Tol.	Active	Estimate	Used
Residue off gas Rate	---	1808 lbmole/hr	---	0.0100 *	2.205 lbmole/hr *	Off	On	Off
Rich oil Rate	---	2173 lbmole/hr	---	0.0100 *	2.205 lbmole/hr *	Off	On	Off
Temperature	90.00 F *	90.00 F	-5.342e-007	0.0100 *	1.800 F *	On	On	On

SPECS

Column Specification Parameters

Residue off gas Rate

Fixed / Ranged: Fixed	Primary / Alternate: Primary	Lower Bound:	---	Upper Bound:	---
Stream: Residue off gas	Flow Basis: Molar				

Rich oil Rate

Fixed / Ranged: Fixed	Primary / Alternate: Primary	Lower Bound:	---	Upper Bound:	---
Stream: Rich oil	Flow Basis: Molar				

Temperature

Fixed / Ranged: Fixed	Primary / Alternate: Primary	Lower Bound:	---	Upper Bound:	---
Stage: 5__TS-1					

SUBCOOLING

Degrees of Subcooling		
Subcool to		

User Variables

PROFILES

General Parameters

Sub-Flow Sheet:	Multicomponent-Absorber (COL1)	Number of Stages:	5 *

Profile Estimates

	Temperature (F)	Net Liquid (lbmole/hr)	Net Vapour (lbmole/hr)
5__TS-1	90.00	2099	1808
4__TS-1	91.30	2118	1902
3__TS-1	92.43	2132	1920
2__TS-1	93.65	2150	1935
1__TS-1	93.90	2173	1952

EFFICIENCIES

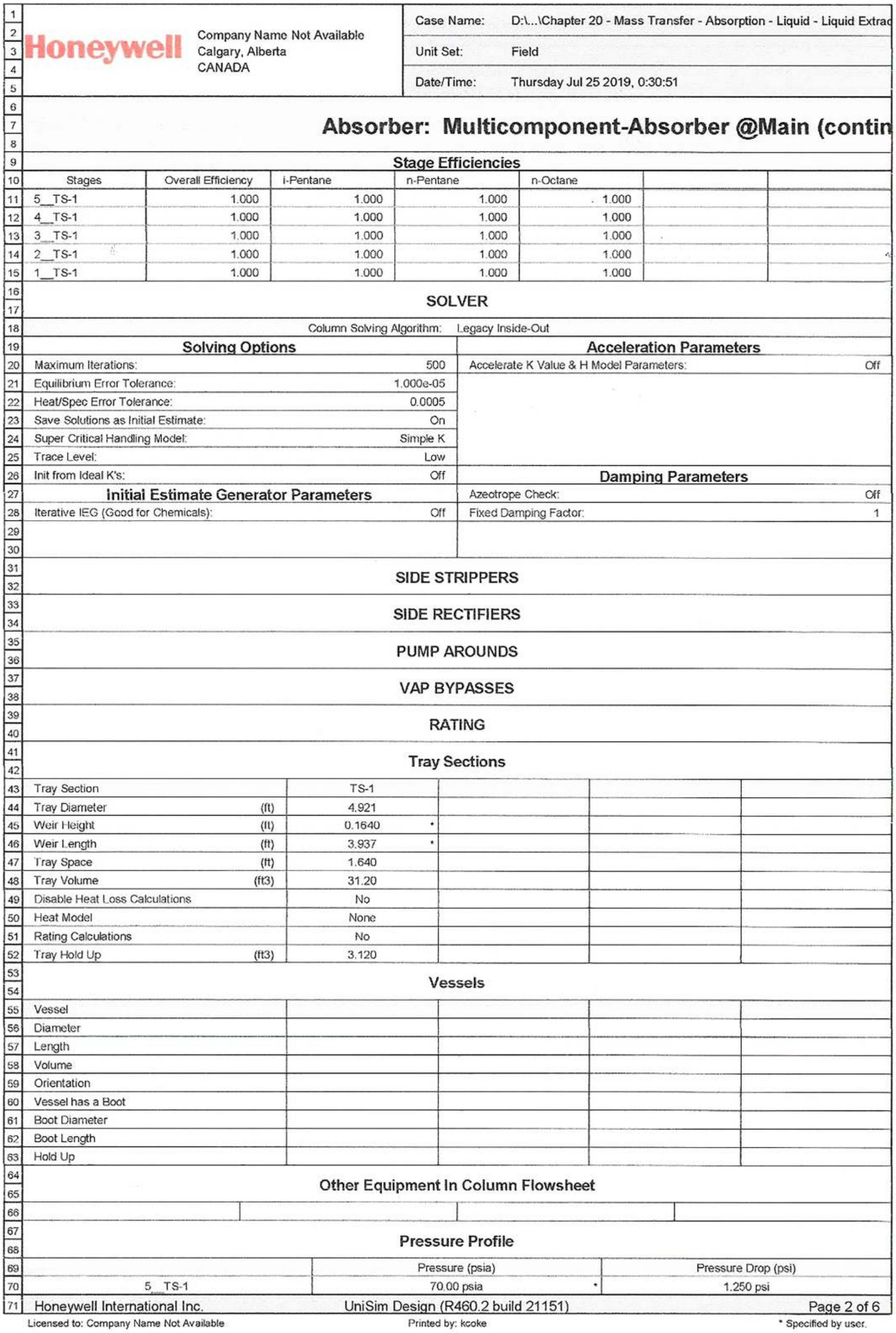
Stage Efficiencies

Stages	Overall Efficiency	Methane	Ethane	Propane	i-Butane	n-Butane
5__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
4__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
3__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
2__TS-1	1.000	1.000	1.000	1.000	1.000	1.000
1__TS-1	1.000	1.000	1.000	1.000	1.000	1.000

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 1 of 6

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

Table 20.52 Simulation results of multicomponent absorber. (*Continued*)



Honeywell	Company Name Not Available Calgary, Alberta CANADA	Case Name:	D:\...\Chapter 20 - Mass Transfer - Absorption - Liquid - Liquid Extrac
		Unit Set:	Field
		Date/Time:	Thursday Jul 25 2019, 0:30:51

Absorber: Multicomponent-Absorber @Main (contin

Stage Efficiencies

Stages	Overall Efficiency	i-Pentane	n-Pentane	n-Octane		
5__TS-1	1.000	1.000	1.000	1.000		
4__TS-1	1.000	1.000	1.000	1.000		
3__TS-1	1.000	1.000	1.000	1.000		
2__TS-1	1.000	1.000	1.000	1.000		
1__TS-1	1.000	1.000	1.000	1.000		

SOLVER

Column Solving Algorithm: Legacy Inside-Out

Solving Options

Maximum Iterations:	500
Equilibrium Error Tolerance:	1.000e-05
Heat/Spec Error Tolerance:	0.0005
Save Solutions as Initial Estimate:	On
Super Critical Handling Model:	Simple K
Trace Level:	Low
Init from Ideal K's:	Off

Acceleration Parameters

Accelerate K Value & H Model Parameters:	Off

Initial Estimate Generator Parameters

Iterative IEG (Good for Chemicals):	Off

Damping Parameters

Azeotrope Check:	Off
Fixed Damping Factor:	1

SIDE STRIPPERS

SIDE RECTIFIERS

PUMP AROUNDS

VAP BYPASSES

RATING

Tray Sections

Tray Section		TS-1			
Tray Diameter	(ft)	4.921			
Weir Height	(ft)	0.1640 *			
Weir Length	(ft)	3.937 *			
Tray Space	(ft)	1.640			
Tray Volume	(ft3)	31.20			
Disable Heat Loss Calculations		No			
Heat Model		None			
Rating Calculations		No			
Tray Hold Up	(ft3)	3.120			

Vessels

Vessel				
Diameter				
Length				
Volume				
Orientation				
Vessel has a Boot				
Boot Diameter				
Boot Length				
Hold Up				

Other Equipment In Column Flowsheet

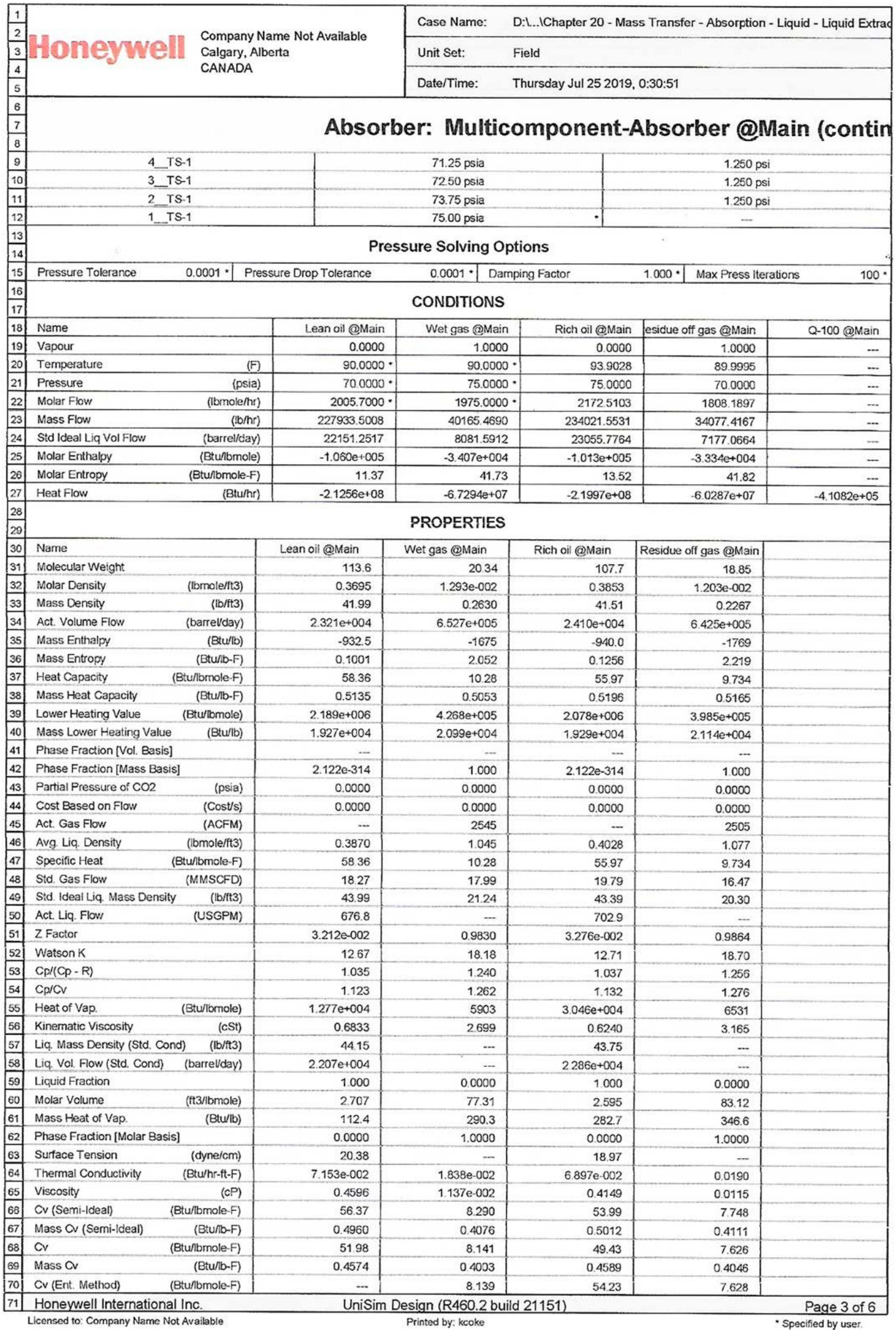
Pressure Profile

	Pressure (psia)	Pressure Drop (psi)
5__TS-1	70.00 psia *	1.250 psi

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 2 of 6

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

Table 20.53 Simulation results of multicomponent absorber. (*Continued*)



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	D:\...\Chapter 20 - Mass Transfer - Absorption - Liquid - Liquid Extrac
Unit Set:	Field
Date/Time:	Thursday Jul 25 2019, 0:30:51

Absorber: Multicomponent-Absorber @Main (contin

4__TS-1	71.25 psia	1.250 psi
3__TS-1	72.50 psia	1.250 psi
2__TS-1	73.75 psia	1.250 psi
1__TS-1	75.00 psia *	---

Pressure Solving Options

Pressure Tolerance	0.0001 *	Pressure Drop Tolerance	0.0001 *	Damping Factor	1.000 *	Max Press Iterations	100 *

CONDITIONS

Name		Lean oil @Main	Wet gas @Main	Rich oil @Main	esidue off gas @Main	Q-100 @Main
Vapour		0.0000	1.0000	0.0000	1.0000	---
Temperature	(F)	90.0000 *	90.0000 *	93.9028	89.9995	---
Pressure	(psia)	70.0000 *	75.0000 *	75.0000	70.0000	---
Molar Flow	(lbmole/hr)	2005.7000 *	1975.0000 *	2172.5103	1808.1897	---
Mass Flow	(lb/hr)	227933.5008	40165.4690	234021.5531	34077.4167	---
Std Ideal Liq Vol Flow	(barrel/day)	22151.2517	8081.5912	23055.7764	7177.0664	---
Molar Enthalpy	(Btu/lbmole)	-1.060e+005	-3.407e+004	-1.013e+005	-3.334e+004	---
Molar Entropy	(Btu/lbmole-F)	11.37	41.73	13.52	41.82	---
Heat Flow	(Btu/hr)	-2.1256e+08	-6.7294e+07	-2.1997e+08	-6.0287e+07	-4.1082e+05

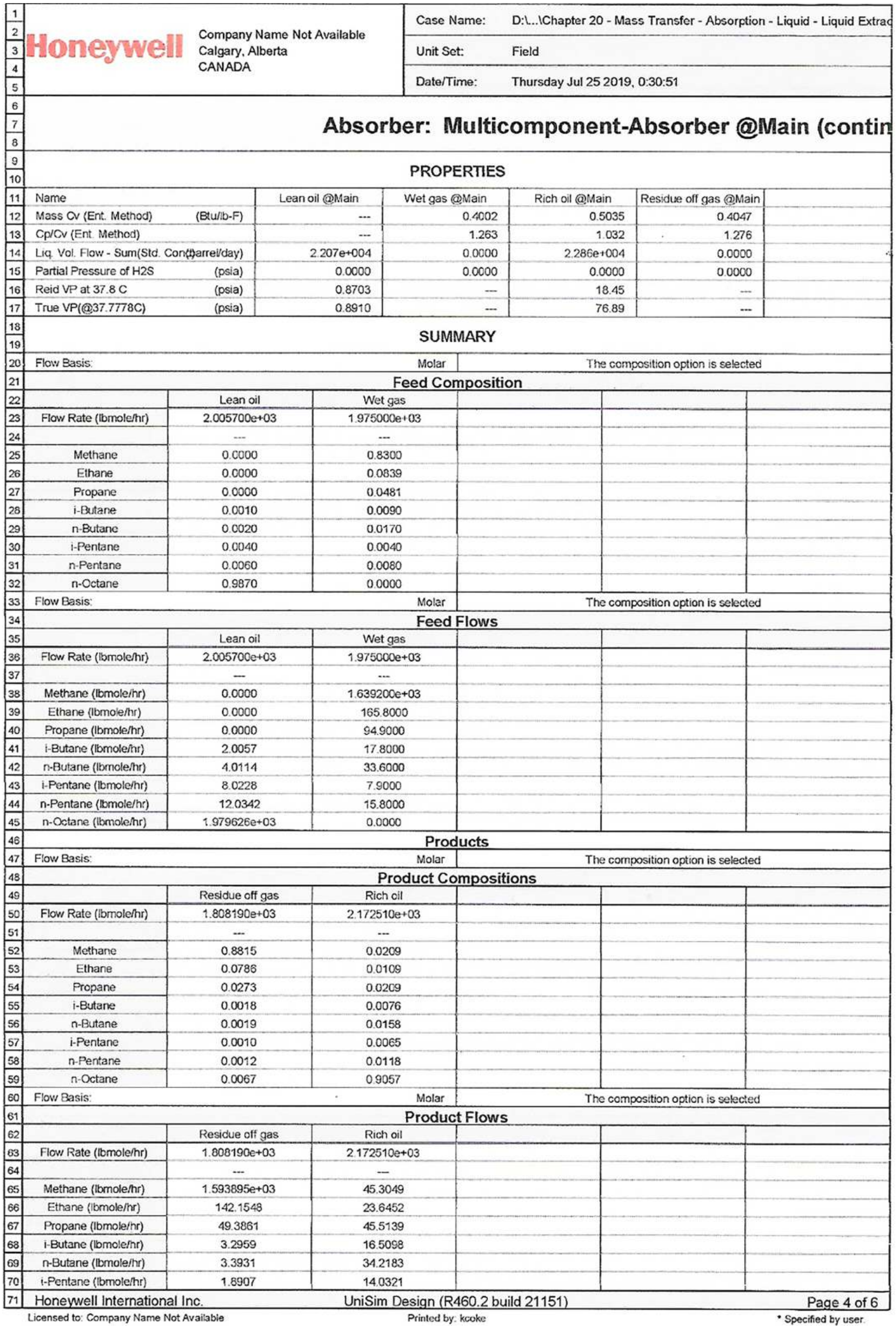
PROPERTIES

Name		Lean oil @Main	Wet gas @Main	Rich oil @Main	Residue off gas @Main	
Molecular Weight		113.6	20.34	107.7	18.85	
Molar Density	(lbmole/ft3)	0.3695	1.293e-002	0.3853	1.203e-002	
Mass Density	(lb/ft3)	41.99	0.2630	41.51	0.2267	
Act. Volume Flow	(barrel/day)	2.321e+004	6.527e+005	2.410e+004	6.425e+005	
Mass Enthalpy	(Btu/lb)	-932.5	-1675	-940.0	-1769	
Mass Entropy	(Btu/lb-F)	0.1001	2.052	0.1256	2.219	
Heat Capacity	(Btu/lbmole-F)	58.36	10.28	55.97	9.734	
Mass Heat Capacity	(Btu/lb-F)	0.5135	0.5053	0.5196	0.5165	
Lower Heating Value	(Btu/lbmole)	2.189e+006	4.268e+005	2.078e+006	3.985e+005	
Mass Lower Heating Value	(Btu/lb)	1.927e+004	2.099e+004	1.929e+004	2.114e+004	
Phase Fraction [Vol. Basis]		---	---	---	---	
Phase Fraction [Mass Basis]		2.122e-314	1.000	2.122e-314	1.000	
Partial Pressure of CO2	(psia)	0.0000	0.0000	0.0000	0.0000	
Cost Based on Flow	(Cost/s)	0.0000	0.0000	0.0000	0.0000	
Act. Gas Flow	(ACFM)	---	2545	---	2505	
Avg. Liq. Density	(lbmole/ft3)	0.3870	1.045	0.4028	1.077	
Specific Heat	(Btu/lbmole-F)	58.36	10.28	55.97	9.734	
Std. Gas Flow	(MMSCFD)	18.27	17.99	19.79	16.47	
Std. Ideal Liq. Mass Density	(lb/ft3)	43.99	21.24	43.39	20.30	
Act. Liq. Flow	(USGPM)	676.8	---	702.9	---	
Z Factor		3.212e-002	0.9830	3.276e-002	0.9864	
Watson K		12.67	18.18	12.71	18.70	
Cp/(Cp - R)		1.035	1.240	1.037	1.256	
Cp/Cv		1.123	1.262	1.132	1.276	
Heat of Vap.	(Btu/lbmole)	1.277e+004	5903	3.046e+004	6531	
Kinematic Viscosity	(cSt)	0.6833	2.699	0.6240	3.165	
Liq. Mass Density (Std. Cond)	(lb/ft3)	44.15	---	43.75	---	
Liq. Vol. Flow (Std. Cond)	(barrel/day)	2.207e+004	---	2.286e+004	---	
Liquid Fraction		1.000	0.0000	1.000	0.0000	
Molar Volume	(ft3/lbmole)	2.707	77.31	2.595	83.12	
Mass Heat of Vap.	(Btu/lb)	112.4	290.3	282.7	346.6	
Phase Fraction [Molar Basis]		0.0000	1.0000	0.0000	1.0000	
Surface Tension	(dyne/cm)	20.38	---	18.97	---	
Thermal Conductivity	(Btu/hr-ft-F)	7.153e-002	1.838e-002	6.897e-002	0.0190	
Viscosity	(cP)	0.4596	1.137e-002	0.4149	0.0115	
Cv (Semi-Ideal)	(Btu/lbmole-F)	56.37	8.290	53.99	7.748	
Mass Cv (Semi-Ideal)	(Btu/lb-F)	0.4960	0.4076	0.5012	0.4111	
Cv	(Btu/lbmole-F)	51.98	8.141	49.43	7.626	
Mass Cv	(Btu/lb-F)	0.4574	0.4003	0.4589	0.4046	
Cv (Ent. Method)	(Btu/lbmole-F)	---	8.139	54.23	7.628	

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 3 of 6

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

Table 20.54 Simulation results of multicomponent absorber. (*Continued*)



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	D:\...\Chapter 20 - Mass Transfer - Absorption - Liquid - Liquid Extrac
Unit Set:	Field
Date/Time:	Thursday Jul 25 2019, 0:30:51

Absorber: Multicomponent-Absorber @Main (contin

PROPERTIES

Name		Lean oil @Main	Wet gas @Main	Rich oil @Main	Residue off gas @Main
Mass Cv (Ent. Method)	(Btu/lb-F)	---	0.4002	0.5035	0.4047
Cp/Cv (Ent. Method)		---	1.263	1.032	1.276
Liq. Vol. Flow - Sum(Std. Con(barrel/day)		2.207e+004	0.0000	2.286e+004	0.0000
Partial Pressure of H2S	(psia)	0.0000	0.0000	0.0000	0.0000
Reid VP at 37.8 C	(psia)	0.8703	---	18.45	---
True VP(@37.7778C)	(psia)	0.8910	---	76.89	---

SUMMARY

Flow Basis: Molar The composition option is selected

Feed Composition

	Lean oil	Wet gas
Flow Rate (lbmole/hr)	2.005700e+03	1.975000e+03
	---	---
Methane	0.0000	0.8300
Ethane	0.0000	0.0839
Propane	0.0000	0.0481
i-Butane	0.0010	0.0090
n-Butane	0.0020	0.0170
i-Pentane	0.0040	0.0040
n-Pentane	0.0060	0.0080
n-Octane	0.9870	0.0000

Flow Basis: Molar The composition option is selected

Feed Flows

	Lean oil	Wet gas
Flow Rate (lbmole/hr)	2.005700e+03	1.975000e+03
	---	---
Methane (lbmole/hr)	0.0000	1.639200e+03
Ethane (lbmole/hr)	0.0000	165.8000
Propane (lbmole/hr)	0.0000	94.9000
i-Butane (lbmole/hr)	2.0057	17.8000
n-Butane (lbmole/hr)	4.0114	33.6000
i-Pentane (lbmole/hr)	8.0228	7.9000
n-Pentane (lbmole/hr)	12.0342	15.8000
n-Octane (lbmole/hr)	1.979626e+03	0.0000

Products

Flow Basis: Molar The composition option is selected

Product Compositions

	Residue off gas	Rich oil
Flow Rate (lbmole/hr)	1.808190e+03	2.172510e+03
	---	---
Methane	0.8815	0.0209
Ethane	0.0786	0.0109
Propane	0.0273	0.0209
i-Butane	0.0018	0.0076
n-Butane	0.0019	0.0158
i-Pentane	0.0010	0.0065
n-Pentane	0.0012	0.0118
n-Octane	0.0067	0.9057

Flow Basis: Molar The composition option is selected

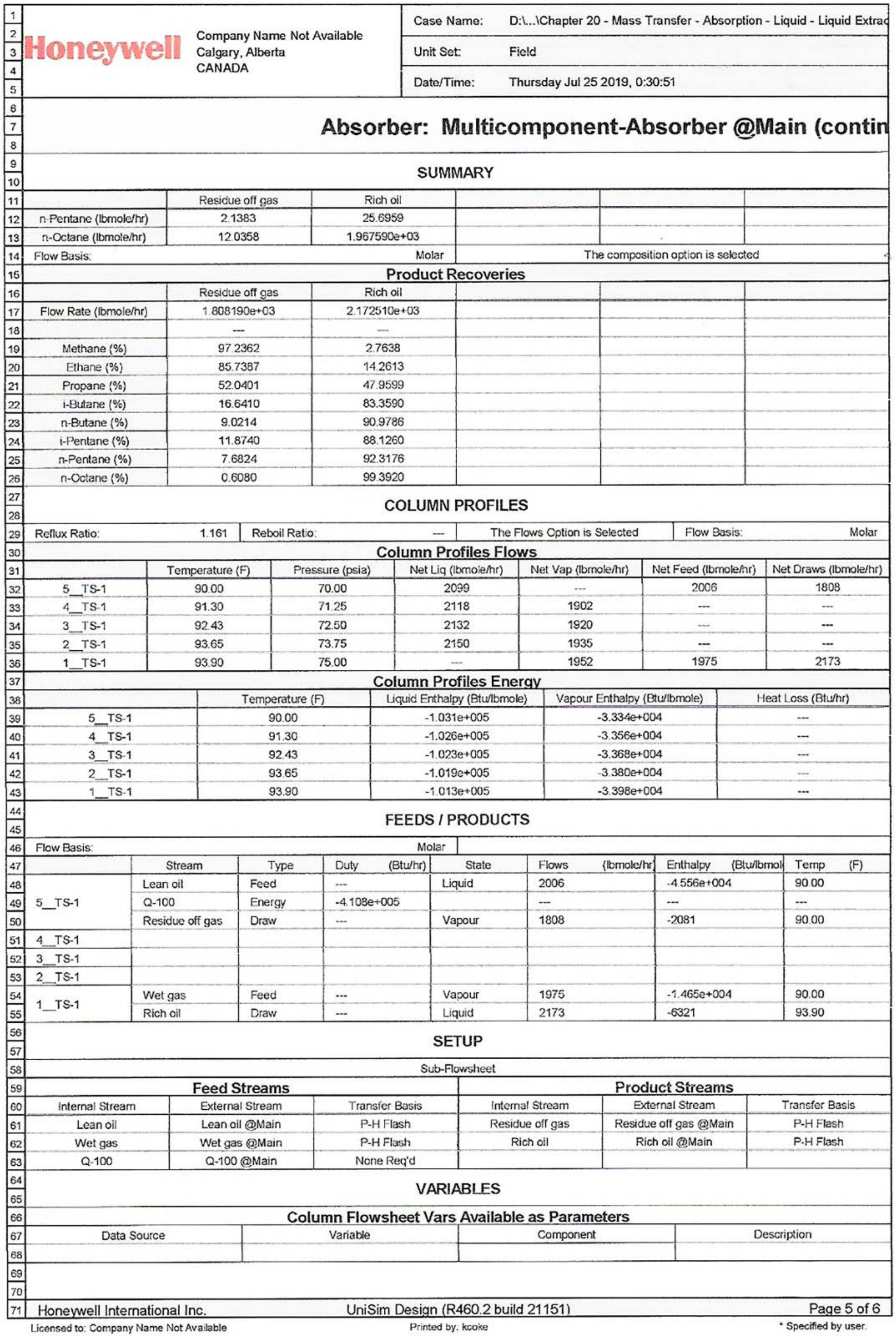
Product Flows

	Residue off gas	Rich oil
Flow Rate (lbmole/hr)	1.808190e+03	2.172510e+03
	---	---
Methane (lbmole/hr)	1.593895e+03	45.3049
Ethane (lbmole/hr)	142.1548	23.6452
Propane (lbmole/hr)	49.3861	45.5139
i-Butane (lbmole/hr)	3.2959	16.5098
n-Butane (lbmole/hr)	3.3931	34.2183
i-Pentane (lbmole/hr)	1.8907	14.0321

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 4 of 6

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

Table 20.55 Simulation results of multicomponent absorber. (*Continued*)



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	D:\...\Chapter 20 - Mass Transfer - Absorption - Liquid - Liquid Extrac
Unit Set:	Field
Date/Time:	Thursday Jul 25 2019, 0:30:51

Absorber: Multicomponent-Absorber @Main (contin

SUMMARY

	Residue off gas	Rich oil			
n-Pentane (lbmole/hr)	2.1383	25.6959			
n-Octane (lbmole/hr)	12.0358	1.967590e+03			
Flow Basis:		Molar	The composition option is selected		

Product Recoveries

	Residue off gas	Rich oil			
Flow Rate (lbmole/hr)	1.808190e+03	2.172510e+03			
	---	---			
Methane (%)	97.2362	2.7638			
Ethane (%)	85.7387	14.2613			
Propane (%)	52.0401	47.9599			
i-Butane (%)	16.6410	83.3590			
n-Butane (%)	9.0214	90.9786			
i-Pentane (%)	11.8740	88.1260			
n-Pentane (%)	7.6824	92.3176			
n-Octane (%)	0.6080	99.3920			

COLUMN PROFILES

Reflux Ratio:	1.161	Reboil Ratio:	---	The Flows Option is Selected	Flow Basis: Molar

Column Profiles Flows

	Temperature (F)	Pressure (psia)	Net Liq (lbmole/hr)	Net Vap (lbmole/hr)	Net Feed (lbmole/hr)	Net Draws (lbmole/hr)
5__TS-1	90.00	70.00	2099	---	2006	1808
4__TS-1	91.30	71.25	2118	1902	---	---
3__TS-1	92.43	72.50	2132	1920	---	---
2__TS-1	93.65	73.75	2150	1935	---	---
1__TS-1	93.90	75.00	---	1952	1975	2173

Column Profiles Energy

	Temperature (F)	Liquid Enthalpy (Btu/lbmole)	Vapour Enthalpy (Btu/lbmole)	Heat Loss (Btu/hr)
5__TS-1	90.00	-1.031e+005	-3.334e+004	---
4__TS-1	91.30	-1.026e+005	-3.356e+004	---
3__TS-1	92.43	-1.023e+005	-3.368e+004	---
2__TS-1	93.65	-1.019e+005	-3.380e+004	---
1__TS-1	93.90	-1.013e+005	-3.398e+004	---

FEEDS / PRODUCTS

Flow Basis: Molar

	Stream	Type	Duty (Btu/hr)	State	Flows (lbmole/hr)	Enthalpy (Btu/lbmol	Temp (F)
5__TS-1	Lean oil	Feed	---	Liquid	2006	-4.556e+004	90.00
	Q-100	Energy	-4.108e+005		---	---	---
	Residue off gas	Draw	---	Vapour	1808	-2081	90.00
4__TS-1							
3__TS-1							
2__TS-1							
1__TS-1	Wet gas	Feed	---	Vapour	1975	-1.465e+004	90.00
	Rich oil	Draw	---	Liquid	2173	-6321	93.90

SETUP

Sub-Flowsheet

Feed Streams			Product Streams		
Internal Stream	External Stream	Transfer Basis	Internal Stream	External Stream	Transfer Basis
Lean oil	Lean oil @Main	P-H Flash	Residue off gas	Residue off gas @Main	P-H Flash
Wet gas	Wet gas @Main	P-H Flash	Rich oil	Rich oil @Main	P-H Flash
Q-100	Q-100 @Main	None Req'd			

VARIABLES

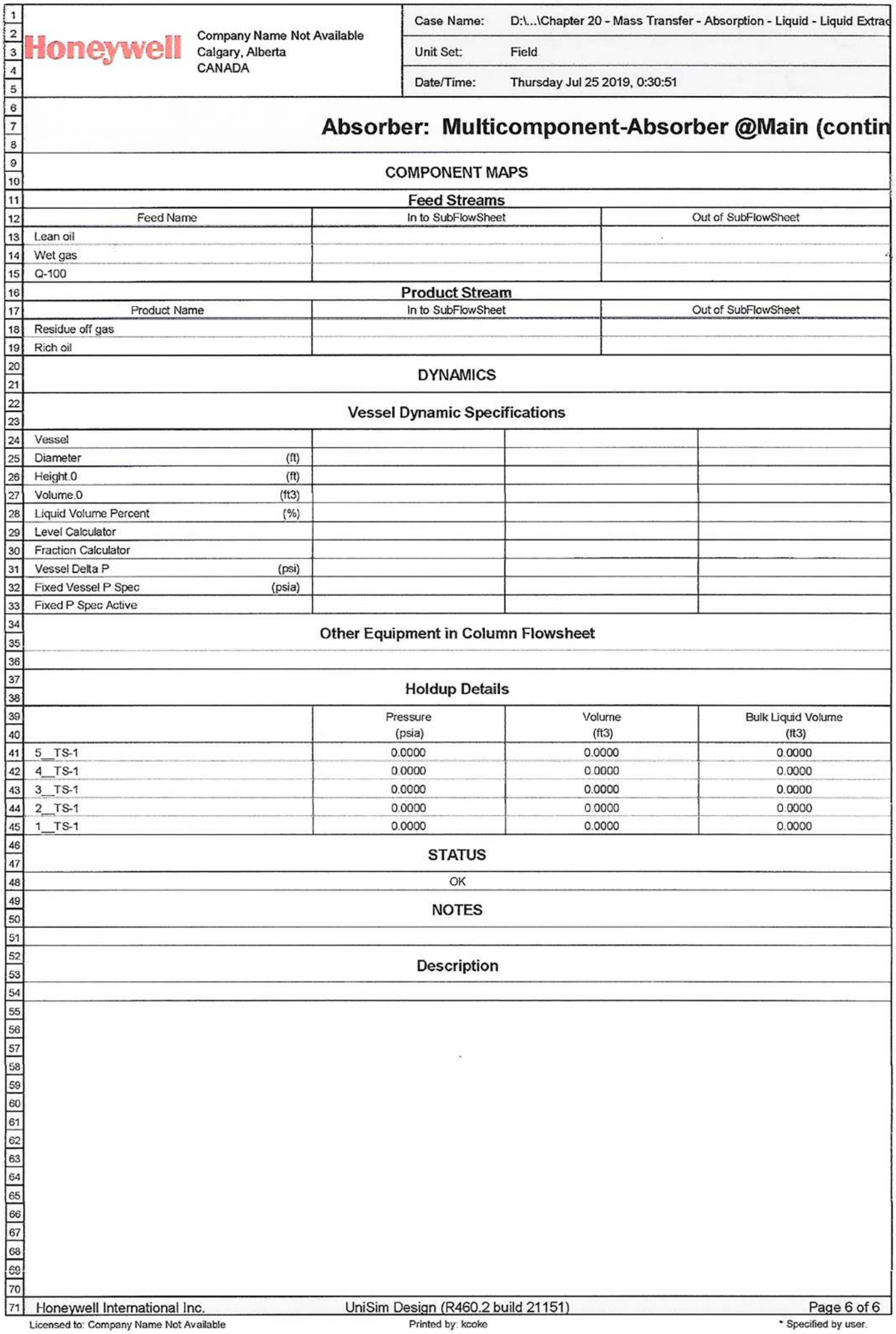
Column Flowsheet Vars Available as Parameters

Data Source	Variable	Component	Description

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 5 of 6

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

Table 20.56 Simulation results of multicomponent absorber. (*Continued*)



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	D:\...\Chapter 20 - Mass Transfer - Absorption - Liquid - Liquid Extrac
Unit Set:	Field
Date/Time:	Thursday Jul 25 2019, 0:30:51

Absorber: Multicomponent-Absorber @Main (contin

COMPONENT MAPS

Feed Streams

Feed Name	In to SubFlowSheet	Out of SubFlowSheet
Lean oil		
Wet gas		
Q-100		

Product Stream

Product Name	In to SubFlowSheet	Out of SubFlowSheet
Residue off gas		
Rich oil		

DYNAMICS

Vessel Dynamic Specifications

Vessel				
Diameter	(ft)			
Height.0	(ft)			
Volume.0	(ft3)			
Liquid Volume Percent	(%)			
Level Calculator				
Fraction Calculator				
Vessel Delta P	(psi)			
Fixed Vessel P Spec	(psia)			
Fixed P Spec Active				

Other Equipment in Column Flowsheet

Holdup Details

	Pressure (psia)	Volume (ft3)	Bulk Liquid Volume (ft3)
5__TS-1	0.0000	0.0000	0.0000
4__TS-1	0.0000	0.0000	0.0000
3__TS-1	0.0000	0.0000	0.0000
2__TS-1	0.0000	0.0000	0.0000
1__TS-1	0.0000	0.0000	0.0000

STATUS

OK

NOTES

Description

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 6 of 6

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

extracted into the second liquid. The two liquid phases are then typically separated, and the extracted component recovered from the solvent phase, e.g., by distillation.

Before the development of solvent extraction processes for uranium and the recovery of metals such as copper, cobalt, and nickel, the industrial use of the technique was found in the petroleum industry. The first commercial process used liquid sulfur dioxide as a selective solvent for aromatic hydrocarbons and demonstrated that solvent extraction could be employed to separate a mixture of hydrocarbons into groups of components of similar chemical type. The original Edeleanu sulfur dioxide process was developed to produce lamp kerosene with clean burning properties and a low aromatic content. While the need for illuminating oils has diminished, there is a continuing requirement for upgraded kerosene for aviation and domestic heating and for lube oils. The removal of the aromatics from the isoparaffins and naphthenes in a lube oil helps to retain its viscous character over a wide range of temperatures. Solvent extraction is not only used to produce high-quality raffinates; it is now employed extensively in the production of high-purity aromatic extracts from catalytic reformates containing some 45–60% aromatics. The benzene (C_6H_6), toluene (C_7H_8), and xylenes (C_8H_{10}) (BTX) produced in this manner are the essential raw material for the petrochemical products such as styrene, nylon, polystyrene, and so on.

The type of material that can be extracted places a constraint on the choice of solvent, which must be satisfied regarding the overall process flexibility and economics. The method chosen for solvent recovery may decide an overall costing; for example, the low boiling point of sulfur dioxide allows recovery by evaporation but at the same time dictates the use of a series of reciprocating compressors. Therefore, sulfur dioxide has been superseded by other solvents where recovery can be more easily achieved by distillation.

20.35.1 BTX Recovery by Solvent Extraction

Commercial processes use Sulfolane, N-methyl pyrrolidone, ethylene glycol, N-formylmorpholine, dimethyl formides (DMF), and dimethylsulfoxide (DMSO). These solvents have a few common characteristics; they all have densities and boiling points, which are higher than the material to be extracted, and all are to a greater or lesser extent polar molecules or water-soluble. Each of these properties has a definite influence on the type of equipment employed and the process flow sheet. For example, a large density difference is desirable for countercurrent flow of the phases and settling in the extraction column, while a high boiling point allows simple separation of the solvent from the extract by distillation. In solvent polarity that strongly influences selectivity for aromatics, it is necessary to add a polar mixing component to the solvent in order to improve its selectivity. The addition of water serves this purpose in the case of dimethylsulfoxide, N-formylmorpholine, and N-methyl pyrrolidone. In general, the two-component solvent has a reduced overall capacity for dissolving hydrocarbons so that when the proportion of aromatics in the feedstock is small, it is disadvantageous to increase the capacity of the solvent by reducing the water content in the mixture and vice versa. Solvents like ethylene glycols are best suited for extraction of light aromatics like benzene, toluene, and xylene from reformed stocks, but not for kerosene. In fact, kerosene quality may not be exposed by smoke point alone; it should be supplemented by flame luminosity. Table 20.57 shows that glycols may not be good solvents for improving smoke point, but the treated stocks possess a good luminosity compared to conventional solvents. While extraction of smoke causing ingredients is desirable, at the same time, retention of illumination like naphthenes is essential. Figure 20.94 illustrates a typical treating process for kerosene, and Table 20.58 shows the commercial solvents for the separation of BTX from light feedstocks.

SnamProgretti's new aromatics extraction process uses N-formylmorpholine solvent. This is quite competitive with present processes, as it has a greater flexibility to feed stocks. Extraction of aromatics by other solvents like DMF, DMSO, and glycols (Figure 20.94) shows that the solvent from the rich phase is separated by water washing and later by distillation. The hydrocarbon phases contain some moisture and have to be dried separately. Steam stripping may not be satisfactory because of the close boiling ranges of solvent and feedstock. The recovery of solvents without encountering hydrolysis has not been reported as this presents a problem. Figures 20.95a–20.95f show operating data of solvents such as DMF, DMSO, and glycols in improving the smoke point. Table 20.59 shows properties of solvents used for elevation of smoke point of kerosene.

Table 20.57 Smoke point and luminosity index.

Improvement in smoke point						
Solvent	**Temperature of extraction, ºC**	**Solvent to feed ratio (S/F)**	**Without dilution (%)**	**With dilution (%)**	**Luminosity Index***	**Calorific value kJ/gm at 30ºC**
DMF	25	2	11	10 (10)%		
	40	2	14	12 (10)%	(32)	46.85
	60	2	12	11 (10%)		
DMSO	30	2	12	13 (5%)	(38)	44.17
	50	2	12	13 (5%)		
	70	2	12	13 (5%)		
Ethylene glycol	30	4	4	4 (5%)	(40)	
	50	4	5	5 (5%)		
Diethylene glycol	30	4	6	6 (5%)	(42)	37.16
	50	4	6	6 (5%)		
Raw kerosene	(smoke point = 18)				(13)	46.85
Tetralin	(smoke point = 9)				(0)	46.85
i-octane	(smoke point = 35)				(50)	47.44

*Luminosity is determined at highest smoke point elevation only.

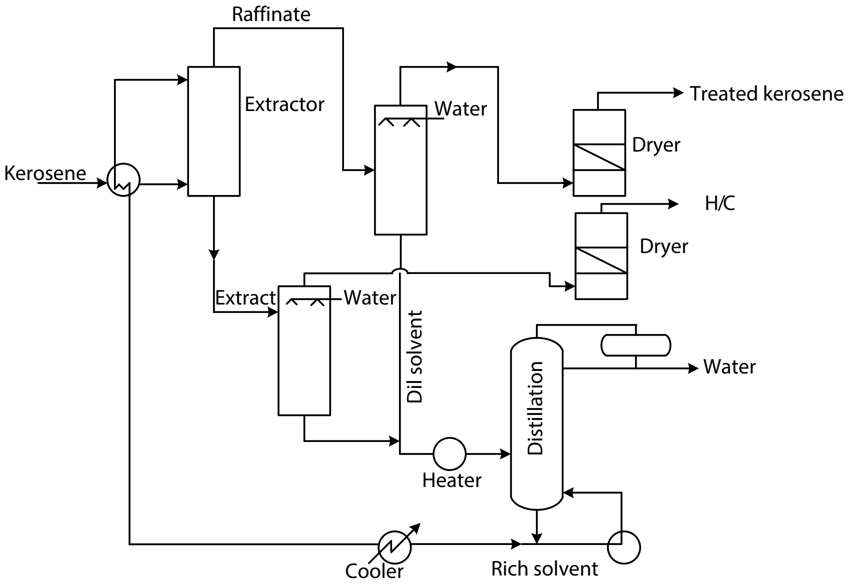


Figure 20.94 Improvement of smoke point with solvents (DMF and DMSO).

Sulfolane, the cyclic compound $(CH_2)_4SO_2$, is a strongly polar molecule, which is very highly selective for aromatics yet possesses a large solvent capacity for hydrocarbons. It was first manufactured on an industrial scale by Shell and together with the rotating disk contactor (RDC); another Shell development forms the basis of the Sulfolane–aromatics process (Figure 20.96). The hydrocarbon feedstock enters at the middle of the extraction column, usually an RDC, where it is treated countercurrently with Sulfolane at a temperature of approximately 120°C. The loaded Sulfolane after leaving the extractor enters an extractive-stripping column where the non-aromatics are removed by extractive distillation together with part of the light aromatics. This stream is recycled to the bottom of the extractor. The Sulfolane and pure aromatics leaving the bottom of the extractive stripper are fed to the recovery column where they are separated by steam distillation under vacuum. The essentially non aromatic raffinate obtained from the top of the column is saturated with Sulfolane, which must be recovered by water washing in another RDC. Ensuring a closed water cycle, the aqueous part of the aromatics steam distillate is used for this purpose; the water washings contaminated with Sulfolane being returned to the aromatics recovery column serves as a stripping steam. Since there are only trace quantities of Sulfolane in the aromatics from the recovery column, it is not necessary to wash this stream, which reduces costs.

In the reforming process, separation methods for recovery of aromatics from hydrocarbon streams employ liquid–liquid solvent extraction to separate the aromatic fraction from other hydrocarbons. The two main extracting solvents are polyglycol and Sulfolane. The polyglycol process is the Udex process developed by Dow Chemical Co. and licensed by UOP, while the Sulfolane process is developed by the Royal Dutch/Shell group and licensed by UOP.

The platformer unit takes in hydrocracked naphtha feed or naphtha from distiller and converts it to produce a single stream of mixed products known as platformate. This is then fed to a Sulfolane extraction unit. The Sulfolane solvent process uses a combination of liquid–liquid extraction and extractive stripping to separate the aromatics from the other hydrocarbons because of the characteristics of the polar solvent. The aromatics and non-aromatics are from the platformer, where aromatics are generated. These are dissolved in Sulfolane extractive unit; the mixture becomes the extract phase (bottom), where the remaining non-aromatics stream forms the raffinate phase (tops).

As the concentration of aromatics in the solvent increases, the solubility of the non-aromatic hydrocarbons in the extract phase also increases. This results in a decrease in selectivity of the solvent and a carryover of some of the non-aromatic hydrocarbons with the extract phase to an extractive stripper. In the stripper, an extractive distillation occurs as the non-aromatic hydrocarbons are stripped from the aromatic–solvent mixture and returned as reflux to the extraction column. The solvent is then recovered from the non-aromatic hydrocarbon free extract stream, leaving an extract that contains less than 1000 ppm (<0.1%) aliphatic. The aromatic recoveries are 99.9, 99.0, and

Table 20.58 Solvents for the separation of Benzene–Toluene and Xylene (BTX) mixtures from light feedstocks.

Solvent	Process	Solvent additives	Operating temperature	Extractor	Solvent recovery
Glycol/water mixtures	UDES process, Universal Oil Products	Solvent can be aqueous diethyl glycol, perhaps with dipropylene glycol or Tetraethylene glycol and water	150°C for diethylene glycol and water	Sieve tray extractor	Distillation
Tetraethylene glycol	Union Carbide Corp.	The solvent is free of water	100°C	Sieve tray extractor	Secondary extraction of BTX into dodecane followed by distillation
Sulfolane	Shell process	None	120°C	Rotating Disc Contactor (RDC) or sieve tray	Distillation
N- Methyl pyrrolidone (NMP)	Arosolvan process, Lurgi	Either water or monoethylene glycol added to improve selectivity	NMP/glycol, 60°C NMP/water, 35°C	Vertical multistage mixer-settler	Distillation
Dimethyl sulfoxide	Institut Francais du Petrole	Solvent contains up to 9% water to improve selectivity	Ambient	Mechanically agitated contactor, e.g., Kuhni column	Aromatics are displaced from DMSO in a second extraction step. Then recovery of secondary paraffinic solvent is by distillation
N-Formly morpholine (FM)	FORMEX process, Snamprogetti	Water addition increases selectivity and also avoids high reboiler temperature during solvent recovery by distillation	40°C	Sieve tray extractor	Usually by distillation, but if benzene content of feedstock is negligible, a secondary extraction with a hydrocarbon solvent is used to remove BTX from the FM
Morpholine derivatives	AROMEX process, Krupp-Koppers	4–5 wt % water content to improve selectivity	Not known	Sieve tray extractor	Distillation

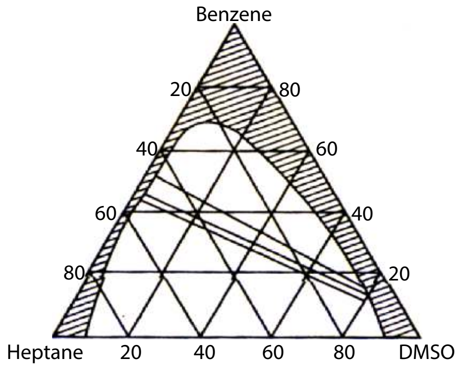


Figure 20.95A Mutual solubility envelope showing tie-lines for extract and raffinate when using undiluted DMSO.

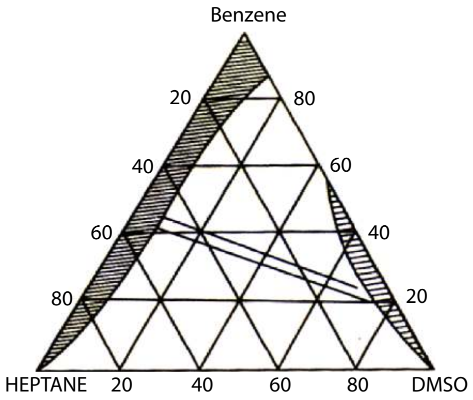


Figure 20.95B The mutual solubility envelope has greater area when DMSO is diluted with 10 wt % water.

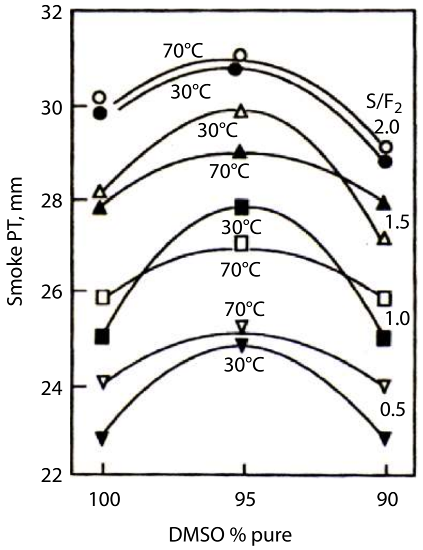


Figure 20.95C Dimethyl sulfoxide as solvent for improvement of smoke point (effect of temperature 30°C and 70°C, S/F ratio and dilution).

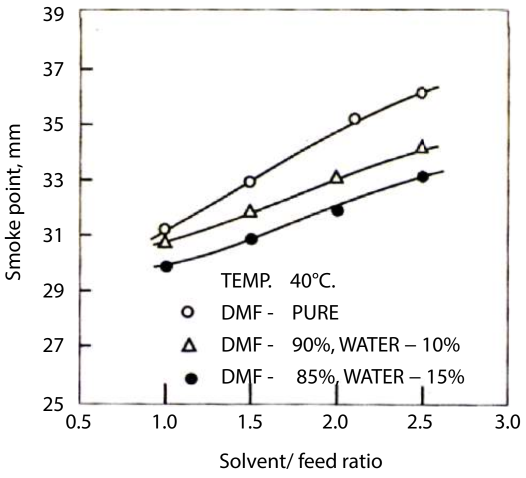


Figure 20.95D Dimethyl formamide as solvent for improvement of smoke point (at 40°C) (effect of S/F ratio, with and without dilution).

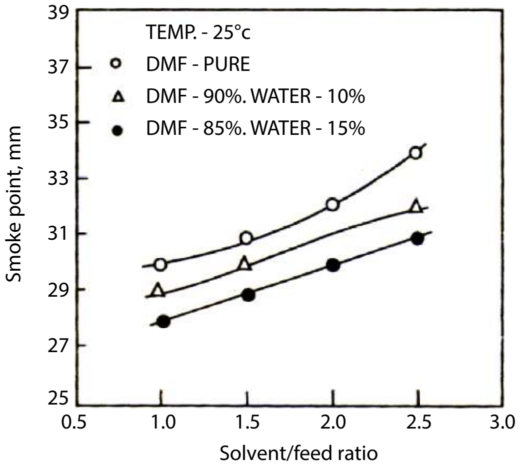


Figure 20.95E Dimethyl Formamide as solvent for improvement of smoke point at 25°C (effect of S/R ratio with and without dilution).

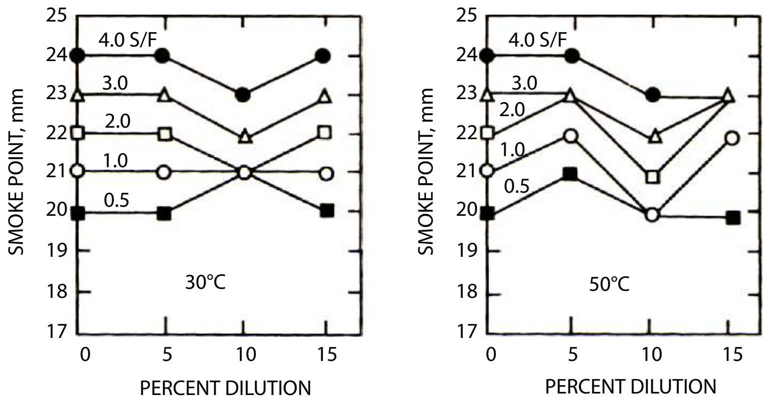


Figure 20.95F Diethylene glycol as solvent for improvement of smoke point (effect of S/F ratio at 30°C and 50°C).

Table 20.59 Properties of solvents used for smoke point improvement.

Solvent	Mol. Wt.	MP, °C	BP, °C	Sp.Gr at 20°C	Sp. heat, J/gm°C	Latent heat, J/gm	Viscosity, cSt at 20°C	Surface tension, dynes 25°C	R.I. at 25°C	Solubility in water wt%
Sulfur dioxide	64	-93	-10	1.45 at -14°C	1.338	387.76 at 17°C	–	–	–	19 at 0°C
Dimethyl formamide	73.1	-61	152.8	0.9489	2.09	575	0.83 at 25°C	35.2	1.4629	Fair
Dimethyl sulfoxide	78.3	18.49	188	1.101	2.05	–	19 at 25°C	–	1.477 at 20°C	Fair
Ethylene glycol	62.068	-14	197	1.116	2.41	450.0	19.0	53	1.43	Fair
Diethylene glycol	106.11	–	245	1.118	2.09	627.0	30.2	48.5	1447 at 20°C	Fair
N-formylmorpholine	115	20	244	1.1528	1.588 at 25°C	435.0	8.13	–	1.4869 at 20°C	Fair

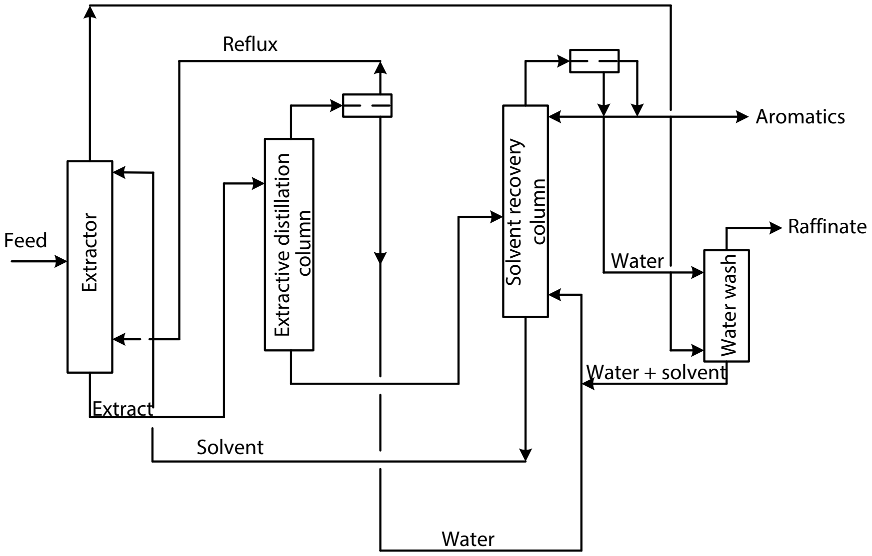


Figure 20.96 Sulfolane aromatics process.

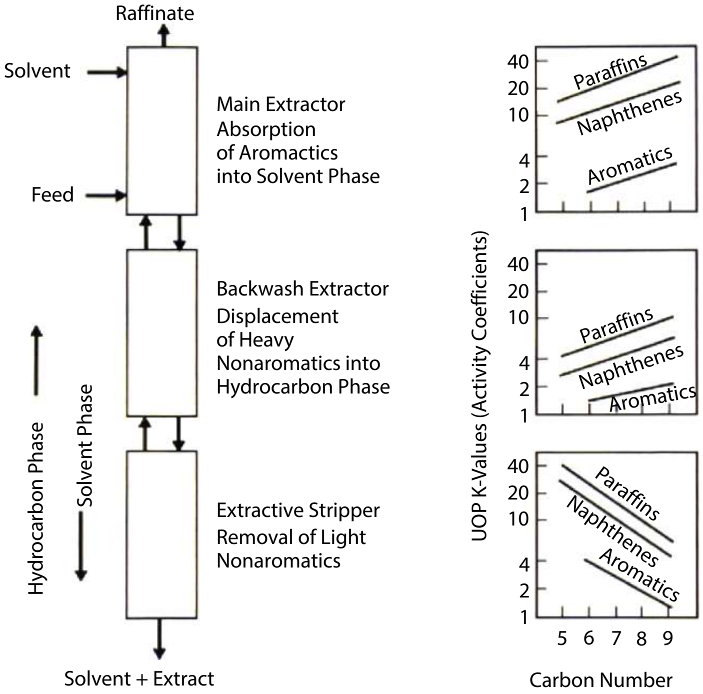


Figure 20.97 Sulfolane process concept (source: Thomas, J. Stoodt and Antoine Negiz, Chapter 2.2. UOP Sulfolane Process, Robert A. Meyers, Handbook of Petroleum Refining Processes, 3rd. ed., McGraw-Hill Handbooks, 2003).

Scale – up Design
Diameter 1778 mm (70 in.)
Height 65 agitated stages
Feed stage 13 from top
Polar solvent feed 0.7: 1.0
Non polar solvent feed: 1.8:1.0
Internals = variable impeller diameters
94% recovery with < 1.0% impurities

Figure 20.98A A photo of liquid–liquid extraction column for aromatic purification (source: Koch Modular Process System (KMPS) LLC).

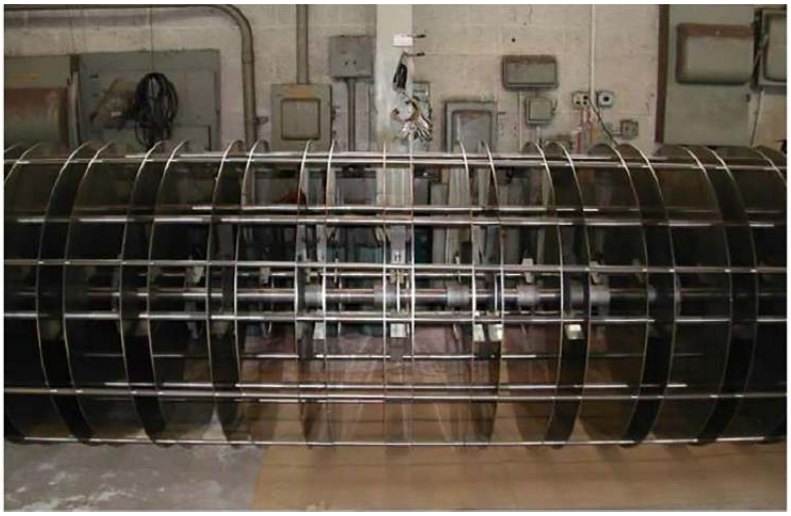


Figure 20.98B Scheibel® Column internal assembly (source: Koch Modular Process System (KMPS) LLC).

97.0% of benzene (C_6H_6), toluene (C_7H_8), and xylene (C_8H_{10}), respectively. Figure 20.97 shows the schematic of the Sulfolane process, and further details are described in volume 1 of the text.

Liquid–liquid extraction is employed for separations when it is either (1) the only method available or (2) the most economical. Because of the need to involve a second separation step to recover the extracted component from the solvent, it is only used for separations, which cannot easily be achieved in a single step. It is a technique that exploits chemical or structural differences between the components of the mixture rather than molecular size and is, therefore, complementary to distillation. Figure 20.98 shows photos of an extraction column for aromatic purification with 65 agitated stages, Scheibel column internal assembly, and KARR® column plate stacked assembly, respectively.

20.36 Process Parameters

Extraction is treated as an equilibrium-stage process and the development of a new extraction process includes the following steps:

Figure 20.98C KARR® Column plate stack assembly (source: Koch Modular Process System (KMPS) LLC).

- Select solvent.
- Obtain physical properties, including phase equilibria.
- Obtain material balance.
- Obtain required equilibrium stages.
- Develop preliminary design contactor.
- Obtain pilot data with preliminary design.
- Determine the effects of scale-up and recycle.
- Obtain final design of system, including the contactor.

The ideal solvent should be easily recovered from the extract and the solvent–solute mixture should have a high relative volatility, low heat of vaporization of the solute, and a high equilibrium distribution coefficient. This translates to a low solvent requirement and a low extract rate fed to the solvent recovery column. These factors should minimize the capital and operating costs associated with the distillation system. Furthermore, the solvent should have a high selectivity (ratio of distribution coefficients), be immiscible with the carrier, have a low viscosity and a high-density difference (compared to the carrier), and have a moderately low interfacial tension. The characteristics of solvents used in extraction applications are discussed later in the chapter.

The main design parameters assigned in an extraction process are

1. Operating temperature
2. Operating pressure
3. Feed flow rate

   a. Composition
   b. Temperature of entering stream
   c. Pressure of entering stream.

As in many separation processes, the temperature and pressure play a significant role in the effectiveness of the separation. For a good split of the feed, the pressure and temperature must be such that all components remain in the liquid phase. The process will be adversely affected if one or more of the components become a vapor, or the extraction may not occur if a large enough portion of the component vaporizes. Additionally, the temperature should be high enough that the components are all soluble with one another. Where the temperatures are very high, finding a suitable solvent for extraction can be difficult. Generally, an extraction process is carried out at ambient temperatures and pressures.

20.36.1 Procedure

In a liquid–liquid extraction unit, a liquid stream (carrier) containing the component (s) to be recovered (solute) is fed into an extractor, where it contains a solvent. The two liquids must be immiscible or only slightly miscible as this allows them to form a dispersion, with one liquid dispersed as droplets in the other. Mass transfer occurs between the droplets (dispersed phase) and the surrounding liquid (continuous phase). For the two liquids to be separated, they must have different densities. The droplets then accumulate above or below the continuous phase, depending on the liquids' relative densities. A boundary is formed between the continuous phase and the droplet dispersion and this is referred to as the interface, which can be at the top or bottom of the extraction column. Figure 20.99 shows the schematic of liquid–liquid extraction. This process is typically carried out in continuous, staged units, which can be operated in either two modes: co-current mixing or countercurrent mixing. The co-current mixing mode is limited to one theoretical stage per extraction unit, while countercurrent mixing is suitable to multiple stages per unit. For this reason, countercurrent mixing is usually preferred to co-current mixing.

Countercurrent extractors can be arranged in one of two ways as shown in Figure 20.100. The choice depends on the density of the solvent relative to that of the solute carrier. If the solvent is less dense than the carrier liquid, the solvent is fed into the bottom of the column, and the solute is carried upward to the top of the extractor, and the carrier liquid is removed from the bottom of the unit (Figure 20.100a). However, if the solvent is more dense than

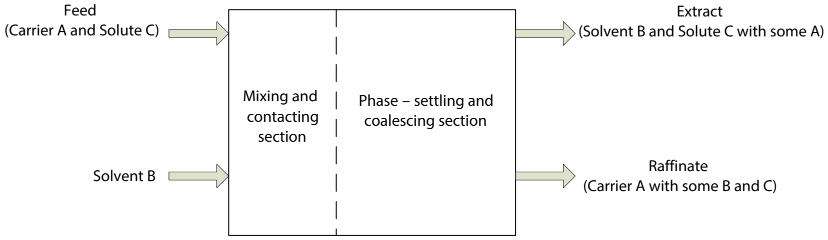


Figure 20.99 An extractor unit has two inlet streams (the liquid carrier containing solute molecules, and the solvent) and two outlet streams (the raffinate and the solute-rich extract).

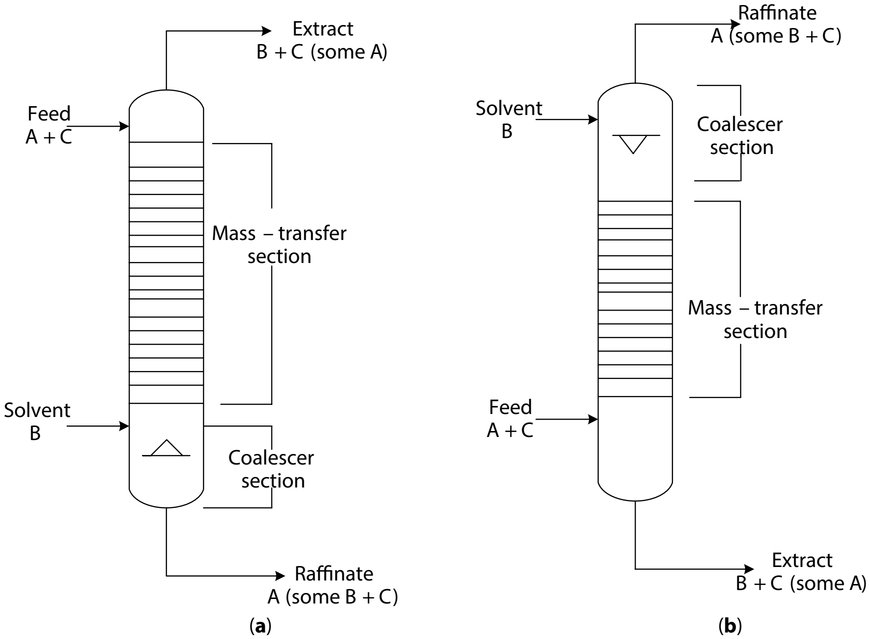


Figure 20.100 Counter-current extraction units can be set up in one of two ways. (a) When the solvent is lighter than the carrier liquid, the solvent is introduced at the bottom of the column and the solute is carried up toward the top of the extractor. (b) When the solvent is heavier than the carrier, the solvent is introduced at the top of the column and the solute is carried downward by the solvent toward the bottom of the column [106].

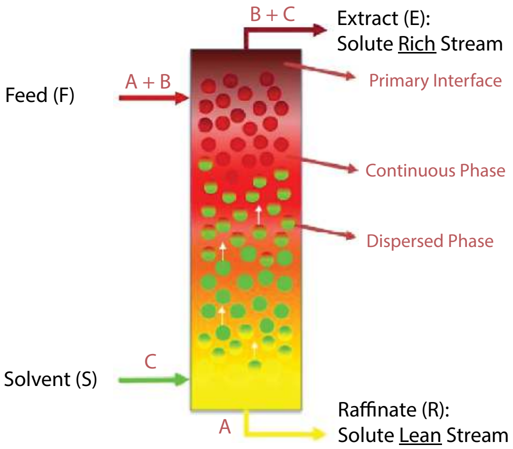


Figure 20.101 Counter-current extraction column (source: Koch Modular Process System (KMPS) LLC).

the carrier liquid, the solvent is fed into the top of the column, the solute is carried downward to the bottom of the extractor, and the carrier stream is removed from the top (Figure 20.100b) [106].

In an extraction column, phases flow countercurrently due to the density difference between the liquids. The column is sized, whether stacked or agitated to generate a certain number of theoretical stages required for efficient transfer of the solute from one liquid phase to the other as illustrated in Figure 20.101.

20.37 Solvents Selection for the Extraction Unit

Solvents used in liquid-liquid extraction are chosen to achieve maximum transfer of the solute from the carrier into the solvent. They must not be completely miscible with the carrier liquid and should have a high affinity for the solute molecules. An ideal solvent should have the following:

1. **Capacity**. This is reflected in the distribution coefficient. It is obviously desirable to use a solvent that could load a high concentration of solute. This will reduce the amount of the solvent in circuit and equipment size and may reduce solvent recovery costs.
2. **Selectivity**. The solvent should be selective for the desired solute over impurities, thus reducing or eliminating the need for scrubbing.
3. **Solvency**. The solvent should have minimal solubility in the raffinate phase. In addition, there should be no tendency for third-phase formation, e.g., due to a low solvency for the solute/extractant complex.
4. **Reversibility**. It must be possible to reverse the extraction process to provide solvent recovery, either by physical or chemical methods. Note this step is often critical economically. If solvent recovery is to be by distillation, a good relative volatility is required. Try to distil over the minority component. If reversal is by chemical conditioning, reagents should be cheap and/or economically recoverable.
5. **Availability**. Ideally, the solvent should be readily available from more than one supplier.
6. **Physical Properties**. A reasonably high interfacial tension promotes phase separation. Similarly, a low viscosity promotes both mass transfer and phase separation and reduces power input requirements. Further properties are:
   a. High boiling point (for ease of material handling).
   b. Low viscosity (for ease of handling).
   c. Nontoxic, nonflammable, and not corrosive to process equipment (for ease of handling).
   d. High solubility for the solute and low solubility for the carrier liquid.
   e. Density difference vs. the carrier liquid greater than 150 kg/m^3.
   f. Mid-level interfacial tension (5–30 dyne/cm).
   g. High resistance to thermal degradation.

   - **h.** Nonreactive with other chemicals involved in the extraction process.

9. **Cost**. The solvent should be cheap. However, this is rarely critical and some of the other factors can have a greater impact on overall economics (e.g., cost of reagents for chemical conditioning).

Viscosity Index Improvement

There is growth of solvent extraction due to high viscosity index lubes and the solvents widely used are furfural, phenol, mixtures of cresols, and propane. The general characteristics responsive to good extraction tendencies for any solvents are:

1. High electronegativity of the hetero atom.
2. High dipole moment.
3. Presence of atoms of O, N, S with activity to form hydrogen bonds.
4. Increased selectivity by substitution in β (5-rings) or m (6-rings) positions.
5. Free rotation of the molecules, no screening by further substitutes.
6. Accumulation of selectively acting groups.

Some of the functional groups are sacrificed, and the final selection of a solvent is based on other factors as [108]:

1. Solubility relationship.
2. Adoptability of various feeds.
3. Ease of recovery.
4. High density (high molecular weight).
5. Less corrosiveness and good stability in the presence of moisture.
6. Low viscosity, low cost, less toxicity.
7. Low interfacial tension.
8. Low freezing point and less latent heat.

Although high selectivity and solvency power govern the extraction process, sometimes highly selective solvents are not chosen in favor of other solvents having less assuming properties. For example, DMF and DMSO are good solvents for the extraction of aromatics but are not favored because of difficulties in their recovery. Phenol and Furfural are competitive solvents as they accomplish their tasks satisfactorily. A survey on the performance of these solvents is presented as follows:

- **a.** Furfural extraction may require a higher temperature of extraction and more number of stages.
- **b.** Yield of raffinate is more with phenol.
- **c.** Phenol treated lubes have better oxidation stability.
- **d.** Viscosity index improvement may be the same in both cases; however, solvent dosage may be slightly more for Furfural on comparative basis.
- **e.** Phenol is more adoptable to different stocks and is unavoidable when naphthenic acids are more.
- **f.** The selectivity of phenol increases with water (up to 10%). This provides a greater advantage over furfural.
- **g.** Furfural is unstable and hence requires nitrogen seal during storage and recovery.
- **h.** Losses can be negligible with phenol.
- **i.** Residual solvent in treated stocks is not harmful provided the solvent is phenol.

After extraction, the pour point of the stock increases between 12 and 15°C. The refineries endeavor to keep the extraction units before dewaxing units resulting in greater loads for the unit. The following dewaxing operation naturally tailors the lubes to the required pour point. Even with many of the above points in favor of phenol, furfural is the preferred solvent because of pollution and toxic nature of phenol [109]. Table 20.60 shows the properties of solvents used for lube treatment.

Table 20.60 Solvents and properties (lube treatment).

Solvent	B.P. °C	M.P. °C	Sp. Gravity at 20°C	Sp. heat J/gm.K	Latent heat J/gm	Solubility in water at 38°C
Furfural	161.6	-36.6	1.162	1.757	419.137	9.0
Phenol	182	41.5	1.07	2.342	445.480	15
Sulfur dioxide	-10	-74	1.45	1.338	387.764 at -17°C	19 at °C
Chlorex	177.8	-51.8	1.22	1.548	260.183	1.02
Duo Sol	180 – 205	–	1.045	2.217	418.3	3.1
Propane	-42	-190	0.51	2.509 (liquid)	422.483	
				1.966 (vapor)		
				1.422		
Nitrobenzene	211	5.4	1.207		296.575	0.1 at 20°C
Aniline	184.4	-6.2	1.027		433.777	

20.38 Phenol Extraction Process of Lubes

Phenol extraction cannot be carried out at a temperature that is less than 50°C because of the high melting point of phenol (40°C). An addition of water up to 10% is approved in all operations, although anhydrous phenol can be used in the extraction. Common operating temperatures range from 60 to 100°C. In practice, a temperature gradient of 10–20°C from the top to bottom is used on the column as this increases the internal turbulence due to mixing of the hot solvent and relatively cold feed [110]. The solvent to feed (S/F) ratio of 1.5–2.5 is used for processing paraffinic and naphthenic stocks. An increase in viscosity index improvement by 25–50 units is confirmed [111]. Some authors use the viscosity gravity constant as a parameter instead of viscosity index [112].

The raffinate yields vary from 50 to 90% depending on the S/F ratio. Higher solvent ratios give better quality of raffinate but fewer yields. Phenol treated oils possess good oxidation stability with less sludge forming and carbon depositing tendencies. In an extraction process, the number of stages plays a vital role in determining the size of the extractor. The number of stages required for phenol treatment is less than four with most units operating at three. The S/F ratio is an essential parameter in extraction operation and for phenol lies in the range of 1.5–2.5 in going from paraffinic to naphthenic oils. The solvent power and its selectivity at lower temperatures match with Chlorex and at a high temperature; it is close to furfural. Therefore, the dual nature of the solvent at all temperatures is essential. Plate or packed columns are generally employed in phenol extraction, and Figures 20.102a–c show the process scheme and operating data, respectively.

20.38.1 Process Description

In Figure 20.102, the charged stock (CS) is heated to 80–100°C by exchanging heat with the raffinate and extract in exchangers (E-1a and E-1b), and then allowed to absorb the phenol vapors in the absorber. These vapors are delivered from the strippers. The stock then flows to the extraction column. This consists of a countercurrent extractor (packed bed or sieve plate column) where a continuous down flow of solvent takes place from the top of the column. The feed being lighter is fed at the bottom of the tower. The extract and raffinate are separated as a distinct layer from the top and are fed into a separator to settle solvent drops. The collected solvent is mingled with the solvent stream that is fed to the extractor. The extract from the bottom product is recycled back to the extractor as a reflux, and the rest is processed for recovering the solvent.

The raffinate phase contains less solvent; hence, it is easily stripped off in the fractionator under vacuum. The stripped raffinate is blown with steam to separate the remaining solvent from the oil phase. If the amount of solvent is small,

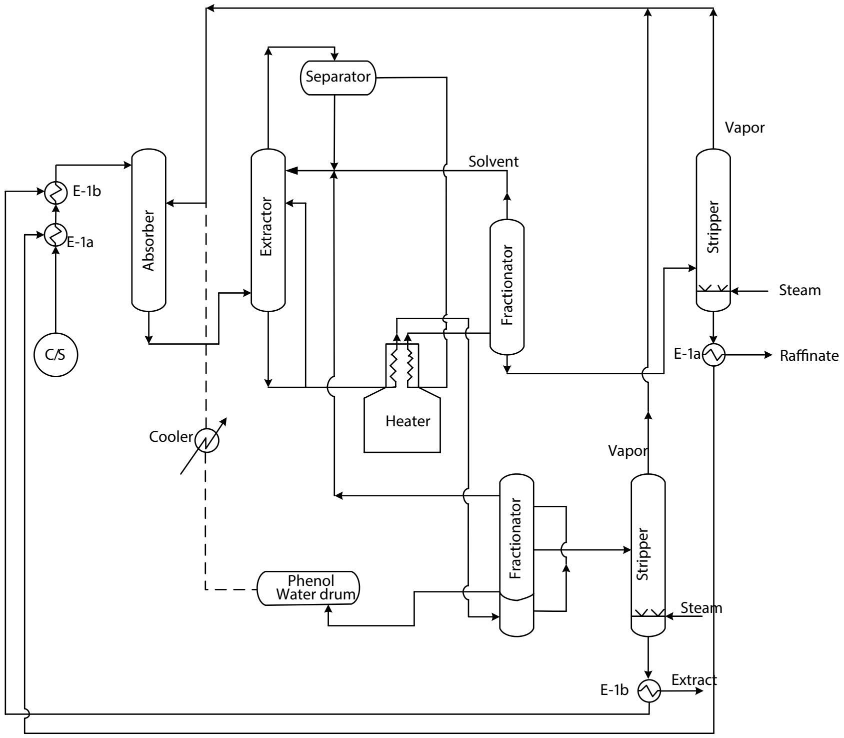


Figure 20.102 Phenol extraction process flow diagram.

then it may be directly steam stripped. Problems may occur while separating the solvent from the extract. This is usually carried out in two stages (i.e., carry out the separation under pressure in lower section and under vacuum in the upper section).

The fractionator is fitted with such an arrangement to vacuum (upper section) while the recovered solvent is fed back to the extractor, and steam stripping of the extract is carried out in a separate unit. The steam blown vapors from the extract and raffinate are absorbed by the fresh charge stock feed and the extraction process is repeated. In general, one liquid is introduced in the continuous phase while the other is in the dispersed phase. The solvent being in large quantity is introduced as the continuous phase. Some of the operating data on phenol extraction are presented in Figures 20.103a and 20.103b, respectively.

20.39 Furfural Extraction Process

Naphthenic stock treatment may not yield convincing results with furfural solvent; however, modern refineries are inclined to use this solvent with the rotating disc contactor (RDC). A temperature range of 75–125°C under a gradient of 20–30°C makes the operation viable. Demands for this solvent are slightly greater, and the number of stages is more in contrast with phenol for the same elevation of viscosity index improvement. Furfural extraction is used not only for lubes but also for gas oils, catalytic cycles oils, and for special applications like butadiene extraction and for the removal of low cetane index components from diesels.

Furfural has a tendency to react with atmospheric oxygen gradually resulting in polymers of furfural; as a consequent upon this change, the solvent power is reduced. To prevent such an unwanted contact, a nitrogen or inert gas seal over the extractor and storage tanks is necessary as also the feedstock should be deaerated before the extraction process starts. The charged stock is heated by passing through heat exchangers HE1 and HE2, through which the raffinate phase and the finished extract pass before reaching storage tanks. This may raise the temperature of the feed between 70 and 80°C. The feed enters the rotating disc contactor at two or three points, and the solvent is fed at the upper section. The raffinate phase being lighter exits from the top of the contactor. This is then heated in a heater H1 and flashed in an evaporator working at a low pressure. This ensures the removal of solvent contaminated with the water vapor if any, and the heavy stream exits at the bottom portion of the evaporator. Steam stripping in the

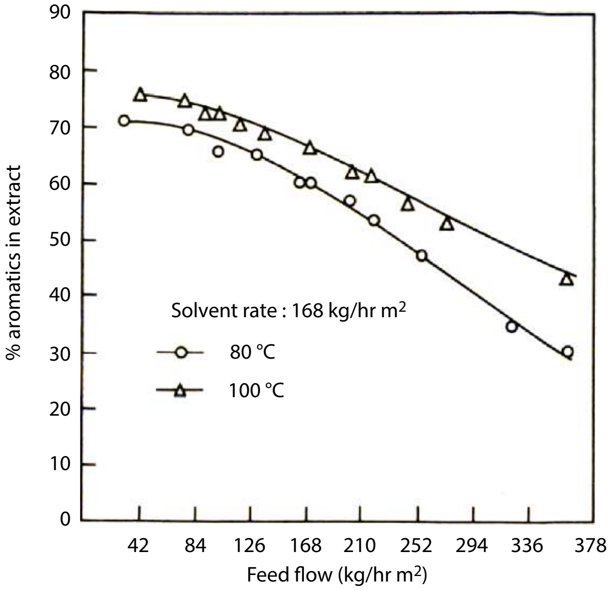


Figure 20.103A Effect of feed rate and temperature operating data on phenol extraction.

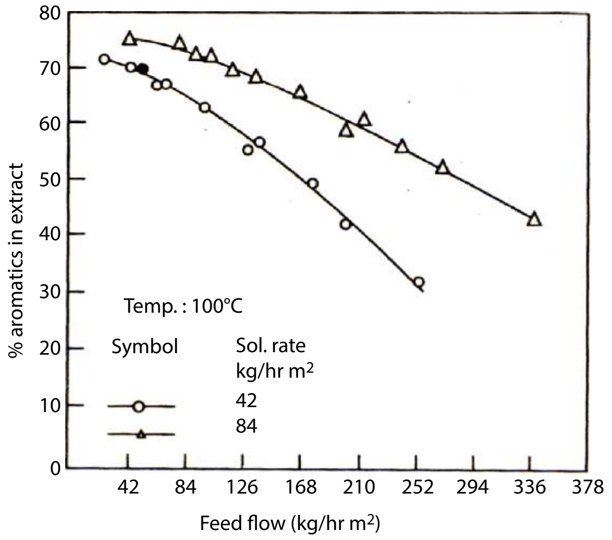


Figure 20.103B Operating data on phenol extraction (effect of feed and solvent rate).

evaporator ensures the removal of the last traces of solvent as solvent vapor–steam mixture. The mixture is cooled in a settler drum.

From the bottom of the evaporator, concentrated raffinate (i.e., free of solvent) is obtained as the finished product and sent for storage through the heat exchanger HE1. The extract from the bottom of the column is heated in a heater H2 and then admitted into the high-pressure zone of a flash tower. The solvent with accompanying water is flashed and later sent for rectification into the furfural column. The heavier portion of the extract is flashed in the low-pressure zone of the flash tower to remove the solvent. An efficient reflux in the tower aids the extraction process. The rich extract is sent to the steam stripper unit to remove the last traces of solvent. The extract is obtained as a finished product from the bottom. Vapors from the strippers (both extract and raffinate) are condensed in a solvent drum to separate furfural and water layers. Furfural water mixture forms an azeotrope (b.pt. 97.7°C) in which 34% of furfural is present by weight. After cooling to room temperature, the furfural-rich layer contains 96% furfural, whereas the water layer contains 8%. The furfural-rich layer is admitted as a reflux into the column. Figure 20.104 illustrates the flow diagram of the extraction process.

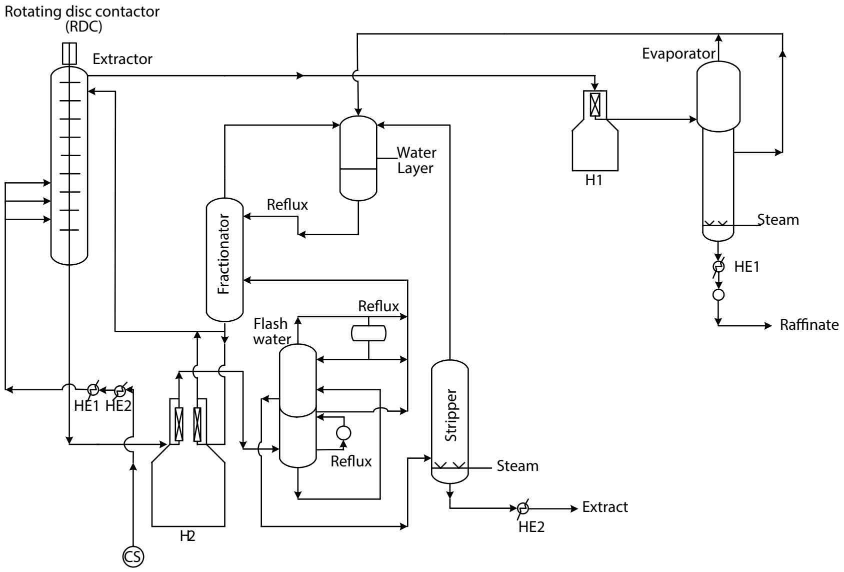


Figure 20.104 Furfural extraction process flow diagram.

20.40 Dispersed-Phase Droplet Size

An extraction column can be designed with either liquid as the dispersed phase: The solvent can be dispersed as droplets in the carrier liquid, or the carrier liquid can be dispersed as droplets in the solvent. In packed columns, it is desirable to disperse the inlet stream with the higher volumetric flow rate into the continuous phase in order to maximize the interfacial area for mass transfer. In agitated columns, the liquid phase with the lower volumetric flow rate is typically dispersed, so that both phases (dispersed and continuous) have similar residence times. Generally, it is desirable to disperse the liquid containing the solute in the extracting solvent, so that the direction of mass transfer is out of the dispersed droplets and into the continuous phase.

The dispersion in the extraction unit is unstable; therefore, the droplets will eventually separate into a bulk phase. Droplet size is a critical parameter in determining the rate at which it occurs. The maximum stable size of droplets in the contacting section of static (nonagitated) extractors depends on the internal energy and physical properties of the fluids and can be determined through statistical measurements taken during tests. Droplets in mechanical agitated extractors will be smaller than those in static extractors. The maximum stable droplet diameter D_m for static is determined by

$$D_m = c\sqrt{\frac{\Pi}{\Delta\rho g}} \tag{20.161}$$

where

c = the proportionality constant
Π = the interfacial tension, dynes/cm
$\Delta\rho$ = difference in density between two fluids, g/cm^3.
g = acceleration due to gravity, 980 cm/s^2.

The direction of mass transfer can influence the behavior of liquid–liquid dispersion, thereby affecting the proportionality constant in Eq. 20.161. The Marangoni effect explains how mass transfer of solute from one phase to the other creates interfacial tension gradients on the surface of the dispersed phase. The transfer of solute from the droplet outward into the continuous phase tends to create larger droplets and higher interfacial area of contact between the two phases. This is represented by a larger proportionality constant. Transfer of solute from the continuous phase into the droplets tends to form smaller droplets and the proportionality constant is smaller. The mass transfer rate across the droplet surface can also influence droplet behavior, with high flux rates causing the droplets to shatter as illustrated in Figures 20.105a and 20.105b.

Mechanically agitated extractors add energy to the system to influence the droplet size and thereby generate enough interfacial area of contact between the two phases. The high shear rate produced by turbulent agitation breaks the droplets into smaller droplets and distorts their shape. Equations have been developed to calculate the maximum drop size for agitated liquid–liquid extraction (107).

Determining the dispersed phase is dependent on

Flow rate (phase ratio)

- For sieve tray and packed columns—disperse the higher flowing phase
- For agitated columns—disperse lower flowing phase.

Viscosity

For efficiency—The disperse is the less viscous phase.

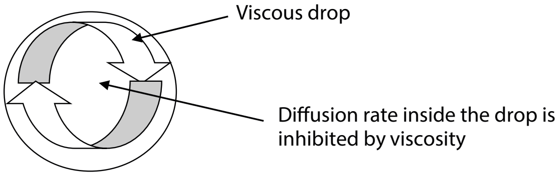


Figure 20.105A The droplet surface.

For capacity—The disperse is the more viscous phase.

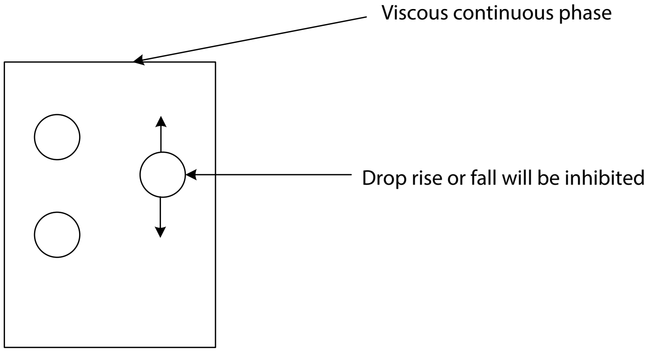


Figure 20.105B Droplet movement.

20.40.1 Surface Wetting

The continuous phase is required to preferentially wet the internals as this minimizes coalescence and therefore maximizes interfacial area (Figure 20.106).

Assume: 30% holdup of disperse phase in 1 m^3 of solution [(Source: Koch Modular Process System (KMPS) LLC)]

Droplet Diameter (m)	**Droplet Volume (m^3)**	**Number of Droplets**	**Droplet SA (m^2)**	**Interfacial Area (m^2/m^3)**
100	0.3	7.16 x 10^{10}	1.26 x 10^{-7}	9022
300	0.3	2.65 x 10^{9}	1.13 x 10^{-6}	2995
500	0.3	5.73 x 10^{8}	3.14 x 10^{-6}	1796

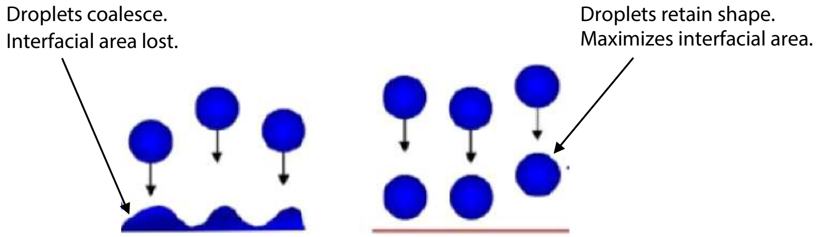


Figure 20.106 Importance of maintaining droplets.

20.40.2 Axial Mixing

Axial (longitudinal) mixing causes the extractor to deviate from ideal plug-flow operation resulting in the dispersed and continuous phases exhibiting a range of residence times. This tends to reduce the concentration driving force for mass transfer in the column. This mixing type can be problematic in large-diameter columns because redistributors have less profound effects in larger columns than in small-diameter ones.

Several mechanisms and conditions that can lead to axial mixing are:

1. Non-uniform droplet size distribution in the contacting section of the column, which can cause it to deviate from plug-flow operation.
2. Chaotic eddy patterns can form when fluids interact with fixed components of the column.
3. Wake shedding can occur when small droplets flow in the wake of larger droplets. Wake shedding can become a problem at extremely high or low volumetric phase ratios.
4. Column walls or other fixed components can exert frictional drag on the liquids and could cause turbulent mixing.

The columns can be designed to minimize axial mixing by dividing the mass transfer section into smaller compartments to minimize deviations from plug flow operation. Sieve plates, for example, act as horizontal barriers that the liquid must pass through establishing a uniform velocity profile downstream of each sieve plate. Structures can be added to the column to divide it into vertical compartments thereby restricting horizontal movement. Figure 20.107 illustrates the extractor flow patterns.

Extractor Flow Patterns

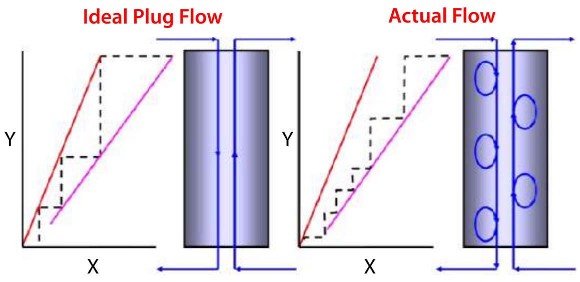


Figure 20.107 The axial or back mixing causes concentration gradients that decrease driving force and therefore increase HETS (source: Koch Modular Process System (KMPS) LLC).

20.40.3 Flooding

The maximum flow rate through an extraction column is limited by flooding. Hydraulic flooding occurs when the two liquid streams can no longer uniformly flow countercurrently pass one another. It is the point at which stable countercurrent extraction is no longer feasible.

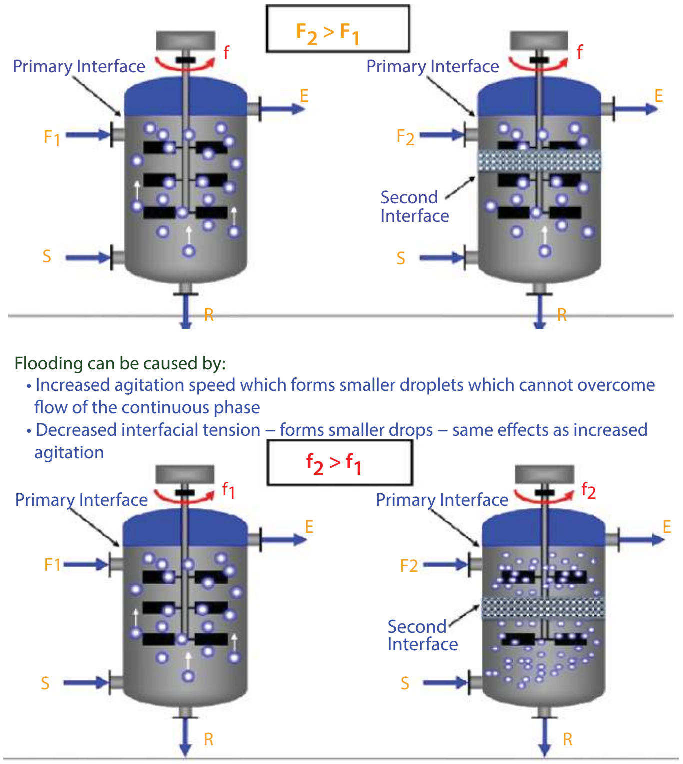
Several instances can result in flooding: excessive hold-up of the dispersed phase droplets; the presence of very small droplets, which significantly increases the mass transfer contacting surface area; and excessive axial mixing.

The dispersed phase holdup can show the presence and degree of flooding. During countercurrent operation, the dispersed phase droplets are held up by the continuous phase that is flowing in the operation direction. The pressure drop over the column is essentially the hydrostatic head of the dispersed and continuous phases. Information from differential-pressure indicators (DPIs) can show the average fluid densities between two DPI tags. Knowing the continuous and dispersed fluid densities can be used to estimate the dispersed phase holdup. The velocity of the dispersed phase droplets relative to the continuous phase is related to the volumetric flowrate of the dispersed and continuous phases. Figure 20.108 shows flooding in a mixer settler.

Flooding is the point where the upward or downward flow of the dispersed phase ceases and a second interface is formed in the column.

Flooding is caused by:

- Increased continuous phase flow rate, which increases drag on droplets.



Flooding can be caused by:
- Increased agitation speed which forms smaller droplets which cannot overcome flow of the continuous phase
- Decreased interfacial tension – forms smaller drops – same effects as increased agitation

Figure 20.108 Flooding in a mixer settler (source: Koch Modular Process System (KMPS) LLC).

20.41 Theory

In liquid–liquid extraction, equilibrium relationship is considered, which is represented by equating the chemical potential of both liquid phases:

$$\mu_i^{LI} = \mu_i^{LII} \tag{20.162}$$

The basic principle of application is that at equilibrium, the activity of each component is the same in both phases. In terms of activity coefficients, this condition is for component i. This relationship reduces to an expression, which is dependent only on the liquid mole fractions and activity coefficients:

$$\gamma_i^{LI} x_i^{LI} = \gamma_i^{LII} x_i^{LII} \tag{20.163}$$

This may be rearranged into a relation of distributions of compositions between the phases,

$$x_i^{LI} = \left(\frac{\gamma_i^{LII}}{\gamma_i^{LI}}\right) x_i^{LII} = K_i\, x_i \tag{20.164}$$

where K_i is the distribution coefficient. The activity coefficients are functions of the compositions of the mixture and the temperature. The distribution coefficient K_i is the ratio of the activity coefficients and is determined from binary infinite dilution coefficient data by

$$K_i = \frac{\gamma_i^{LII,\infty}}{\gamma_i^{LI,\infty}} \tag{20.165}$$

Binary interaction parameters (A_{ij}) and infinite dilution activity coefficients are available for a wide variety of binary pairs. Therefore, the ratio of the solute infinite dilution coefficient in solvent-rich phase to that of the second phase (LII) will provide an estimate of the equilibrium distribution coefficient. The method can provide a reasonable estimate of the distribution coefficient for dilute cases by

$$RT \ln\left[\gamma_{i,j}^{LII,\infty}\right] = A_{i,j}^{LII} \tag{20.166}$$

Rearranging Eq. 20.166 and substituting into Equation 20.165 gives

$$K_i = \frac{\gamma_i^{LII,\infty}}{\gamma_i^{LI,\infty}} = \frac{e^{A_{ij}^{LII}/RT}}{e^{A_{ij}^{LI}/RT}} \tag{20.167}$$

We can use activity coefficient models, such as UNIFAC (UNIquac Functional group Activity Coefficient), UNIQUAC (Universal quasichemical), and NRTL (nonrandom two-liquid) to determine the mole fractions. All three models apply for liquid–liquid equilibrium. For a multicomponent system, the UNIQUAC equation for the liquid phase activity coefficient is represented as

$$\ln \gamma_i = \ln \gamma_i(\text{combinatorial}) + \ln \gamma_i(\text{residual}) \tag{20.168}$$

The combinatorial and residual activities are based on the statistical mechanical theory and allowed the local compositions to result from the size and energy differences between the molecules in the mixture. Table 20.61 shows the NRTL and UNIQUAC correlations for Activity Coefficients of three-component mixtures.

20.42 Nernst's Distribution Law

Nernst's distribution law is important in connection with extraction. A substance is to be separated from a feed. This is referred to as the substance to be extracted. For this purpose, an extractant or solvent is added. The substance to be extracted will distribute itself between the solvent and the rest of the feed, which is referred to as the extract phase and the raffinate phase. According to Nernst's law, the distribution coefficient is

$$K = \frac{C_{\text{substance (solute) in extract phase}}}{C_{\text{substance (solute) in raffinate phase}}} \tag{20.169}$$

where

$C_{\text{substance (solute) in extract phase}}$ = Concentration of substance (solute) to be extracted in extract phase (kg/m^3)
$C_{\text{substance (solute) in raffinate phase}}$ = Concentration of substance (solute) to be extracted in raffinate phase (kg/m^3)
k = distribution coefficient (dimensionless)

Distribution coefficient data for dilute solute concentrations have been compiled by Treybal in Perry's Handbook. A sampling of the data presented by Treybal and additional distribution coefficient data are given in Table 20.62.

Selectivity term is more often used in extraction work. Selectivity away from unity results in better extraction with less difficulty. Solvent power is the capacity of a solvent to dissolve a substance at a given temperature. A good solvent should possess high solvent power and good selectivity.

Conventional procedure of representing the ternary system is carried out in triangular plots. Equilateral triangles serve the purpose most as the apex of the opposite side perpendicular distance remains the same. Each apex is assigned with 100% purity of a component. The concentration at any point inside the triangle can be determined by drawing the perpendiculars to the opposite sides from that point. Percentage concentration can be determined by measuring the height. Binodal curve is the miscible region with immiscible and, therefore, acts as a boundary of the heterogeneous area and bears the tie line points.

With higher heterogeneous region, better extraction is possible due to the flexibility in handling the feeds of variable concentrations. As the temperature of the extraction increases, the area shrinks thus impairing extraction very much. Though higher temperature increases solubility and selectivity, the operational region shrinks very much and thus the flexibility in choosing very high concentrations of the feed diminishes. A technique to increase such area is by adding an anti-solvent (e.g. water), which can open the closed binodal curve to give liberal access to extraction. For an example, in DMSO, benzene, heptane system, heptane is, in fact, a better solvent for removing benzene from DMSO, but DMSO is used to remove benzene from heptane–benzene mixture because the selectivity is away from unity (Figures 20.95a and 20.95b).

20.43 Tie Lines

The equilibrium compositions of extract and raffinate lie in a binodal curve, i.e., the line that joins these two points is known as the tie line. The slopes of the tie lines vary even for the same system, and the distinction disappears between the raffinate and extract at one point, which is known as the plait point. The slope of the tie line gives valuable information regarding the capacity of the solvent. If the slope of the tie line is away from the solvent apex, then the solvent is considered to be very good.

Table 20.61 NRTL and UNIQUAC correlations for activity coefficients of three-component mixtures.

NRTL
$\begin{aligned} \ln \gamma_i = {} & \frac{\tau_{1i} G_{1i} x_1 + \tau_{2i} G_{2i} x_2 + \tau_{3i} G_{3i} x_3}{G_{1i} x_1 + G_{2i} x_2 + G_{3i} x_3} \\ & + \frac{x_1 G_{i1}}{x_1 + G_{12} x_2 + G_{13} x_3} \left[\tau_{i1} - \frac{x_2 \tau_{21} G_{21} + x_3 \tau_{31} G_{31}}{x_1 + x_2 G_{21} + x_3 G_{31}} \right] \\ & + \frac{x_2 G_{i2}}{G_{12} x_1 + x_2 + G_{32} x_3} \left[\tau_{i2} - \frac{x_1 \tau_{12} G_{12} + x_3 \tau_{32} G_{32}}{x_1 G_{12} + x_2 + x_3 G_{32}} \right] \\ & + \frac{x_3 G_{i3}}{G_{13} x_1 + G_{23} x_2 + x_3} \left[\tau_{i3} - \frac{x_1 \tau_{13} G_{13} + x_2 \tau_{23} G_{23}}{x_1 G_{13} + x_2 G_{23} + x_3} \right] \end{aligned}$
$\tau_{ii} = 0$
$G_{ii} = 1$
UNIQUAC
$\begin{aligned} \ln \gamma_i = {} & \ln \frac{\phi_i}{x_i} + 5 q_i \ln \frac{\phi_i}{x_i} + l_i - \frac{\phi_i}{x_i}(x_1 l_1 + x_2 l_2 + x_3 l_3) + q_i [1 - \ln(\theta_1 \tau_{1i} + \theta_2 \tau_{2i} + \theta_3 \tau_{3i})] \\ & - \frac{\theta_1 \tau_{i1}}{\theta_1 + \theta_2 \tau_{21} + \theta_3 \tau_{31}} - \frac{\theta_2 \tau_{i2}}{\theta_1 \tau_{12} + \theta_2 + \theta_3 \tau_{32}} - \frac{\theta_3 \tau_{i3}}{\theta_1 \tau_{13} + \theta_2 \tau_{23} + \theta_3} \end{aligned}$
$\tau_{1i} = 1$
$\phi_i = \frac{r_i x_i}{r_1 x_1 + r_2 x_2 + r_3 x_3}$
$\theta_i = \frac{q_i x_i}{q_1 x_1 + q_2 x_2 + q_3 x_3}$
$l_i = 5(r_i - q_i) - r_1 + 1$

[a]NRTL equation: There is a pair of parameters $\mathbf{g}_{jk}$ and $\mathbf{g}_{kj}$ for each pair of substances in the mixture; for three substances, there are three pairs. The other terms of the equations are related to the basic ones by

$$\tau_{jk} = g_{jk}/RT$$

$$G_{jk} = \exp(-\alpha_{jk}\, \tau_{jk})$$

For liquid–liquid systems usually, $\alpha_{jk} = 0.4$
UNIQUAC equation: There is a pair of parameters $\mathbf{u}_{jk}$ and $\mathbf{u}_{kj}$ for each pair of substances in the mixture:

$$\tau_{jk} = \exp\left(-u_{jk}/RT\right)$$

The terms with single subscripts are properties of the pure materials, which are usually known or can be estimated. The equations are extended readily to more components.

Table 20.62 Distribution coefficient of dilute solute systems.

Component A = Carrier, Component B = Solute, and Component S = Extraction solvent; K is the distribution coefficient (wt/wt).			
Component B	**Component S**	**Temp, °C**	**K**
A = ethylene glycol			
Acetone	Amyl acetate	31	1.838
Acetone	n-Butyl acetate	31	1.940
Acetone	Ethyl acetate	31	1.850
A = Furfural			
Trilinolein	n-Heptane	30	47.5
Triolein	n-Heptane	30	95
A = n-hexane			
Toluene	Sulfolane	25	0.336
Xylene	Sulfolane	25	0.302
A = n-octane			
Toluene	Sulfolane	25	0.345
Xylene	Sulfolane	25	0.245
A = water			
Acetaldehyde	n-Amyl alcohol	18	1.43
Acetaldehyde	Furfural	16	0.967
Acetic acid	1 - Butanol	26.7	1.613
Acetic acid	Cyclohexanol	26.7	1.325
Acetic acid	Di-n-butyl ketone	25-26	0.379
Acetic acid	Diisopropyl carbinol	25-26	0.800
Acetic acid	Ethyl acetate	30	0.907
Acetic acid	Isopropyl ether	20	0.248
Acetic acid	Methyl acetate		1.273
Acetic acid	Methyl cyclohexanone	25-26	0.930
Acetic acid	Methyl isobutyl ketone	25 25-26	0.657 0.755
Acetic acid	Toluene	25	0.0644

(*Continued*)

Table 20.62 Distribution coefficient of dilute solute systems. (*Continued*)

Component A = Carrier, Component B = Solute, and Component S = Extraction solvent; K is the distribution coefficient (wt/wt).			
Component B	**Component S**	**Temp, °C**	**K**
Acetone	n-Butyl acetate		1.127
Acetone	Chloroform	25 25	1.830 1.720
Acetone	Dibutyl ether	25-26	1.941
Acetone	Diethyl ether	30	1.00
Acetone	Ethyl acetate	30	1.500
Acetone	Ethyl butyrate	30	1.278
Acetone	n-Hexane	25	0.343
Acetone	Methyl acetate	30	1.153
Acetone	Methyl isobutyl ketone	25 - 26	1.910
Acetone	Toluene	25 - 26	0.835
Aniline	n-Heptane	25 50	1.425 2.20
Aniline	Methylcyclohexane	25 50	2.05 3.41
Aniline	Toluene	25	12.91
tert-Butanol	Ethyl acetate	20	1.74
Butyric acid	Methyl butyrate	30	6.75
Citric acid	25% Triisooctylamine/Chloroform	25	14.1
Citric acid	25% Triisooctylamine/1-Octanol	25	41.5
p-Cresol	Methylnaphthalene	35	9.89
Ethanol	n- Butanol	20	3.00
Ethanol	Di-n-propyl ketone	25 - 26	0.592
Ethanol	3-Heptanol	25	0.783
Ethanol	n-Hexanol	28	1.000
Ethanol	Sec-Octanol	28	0.825
Ethylene glycol	Furfural	25	0.315
Formic acid	Methylisobutyl carbinol	30	1.218
Furfural	Toluene	25	5.64
Lactic acid	iso-Amyl alcohol	30	0.352
Lactic acid	25% Triisooctylamine/Chloroform	25	19.2

(*Continued*)

Table 20.62 Distribution coefficient of dilute solute systems. (*Continued*)

Component A = Carrier, Component B = Solute, and Component S = Extraction solvent; K is the distribution coefficient (wt/wt).			
Component B	**Component S**	**Temp, °C**	**K**
Lactic acid	25% Triisooctylamine/1-Octanol	25	25.9
Malic acid	25% Triisooctylamine/Chloroform	25	30.7
Malic acid	25% Triisooctylamine/1-Octanol	25	59.0
Methanol	n-Butanol	0	0.600
Methanol	p - Cresol	35	0.313
Methanol	Ethyl acetate	0 20	0.0589 0.238
Methanol	n-Hexanol	28	0.565
Methanol	Phenol	25	1.333
Methyl-n-butyl ketone	n-Butanol	37.8	53.4
Methyl ethyl ketone	Cyclohexane	25	1.775
Methyl ethyl ketone	n-Heptane	25	1.548
Methyl ethyl ketone	n-Hexane	25 37.8	1.775 2.22
Methyl ethyl ketone	1,1,2-Trichloroethane	25	3.44
Methyl ethyl ketone	Trichloroethylene	25	3.27
Methyl ethyl ketone	2,2,4-Trimethylpentane	25	1.572
Oxalic acid	25% Triisooctylamine/Chloroform	25	25.5
Oxalic acid	25% Triisooctylamine/1-Octanol	25	46
Phenol	Ethyl acetate	25	0.048
Phenol	Isoamyl acetate	25	0.046
Phenol	Isopropyl acetate	25	0.040
iso-Propanol	Carbon tetrachloride	20	1.405
iso-propanol	Diisopropyl ether	25	0.406
n- Propanol	Iso-Amyl alcohol	25	3.34
n- Propanol	n-Butanol	37.8	3.61
n- Propanol	Ethyl acetate	0 20	1.419 1.542
n- Propanol	n- Heptane	37.8	0.540
n- Propanol	n- Propyl acetate	20	1.55
Propionic acid	Ethyl acetate	30	2.77

(*Continued*)

Table 20.62 Distribution coefficient of dilute solute systems. (*Continued*)

Component A = Carrier, Component B = Solute, and Component S = Extraction solvent; K is the distribution coefficient (wt/wt).			
Component B	**Component S**	**Temp, °C**	**K**
Propionic acid	Ethyl butyrate	36	1.470
Propionic acid	Ethyl propionate	28	0.510
Propionic acid	Methyl butyrate	30	2.15
Propionic acid	Methylisobutyl carbinol	30	3.52
Propionic acid	Monocholorobenzene	30	0.513
Propionic acid	Toluene	31	0.515
Propionic acid	Tricholoroethylene	30	0.496
Pyridine	Monocholorobenzene	25	2.10
Pyridine	Toluene	25	1.900
Pyridine	Xylene	25	1.260
Sodium chloride	1-Methyldodecyl amine	30	0.693
Sodium chloride	1-Methyloctyl amine	30	0.589
A = Salt water			
Citric acid	2-Butanol	25	0.534

The tie line concentration can be plotted on the x–y diagram as in the McCabe Thiele equilibrium diagram, which simplifies the evaluation of the number of stages involved in the operation. Minimum solvent required in the operation can be obtained by extending the tie lines; the larger the distance between the meeting point of the highest tie line and the lowest tie line from the solvent apex, the higher the solvent required.

When extraction is conducted in conventional packed bed extractors, the column diameter and packing size are important parameters. Packing size to column size should be as large as possible (minimum 8) to avoid channeling. The minimum size of packing required for a system is given by the critical packing size defined by [113]

$$d_{FC} = 2.42\left[\frac{\gamma}{g\Delta\rho}\right]^{0.5} \qquad (20.170)$$

where

γ = surface tension of the system at boundary, dynes/cm
$\Delta\rho$ = difference in density of phases, gm/cm^3.
g = acceleration due to gravity, 980 cm/s^2

Example 20.12

In a phenol extraction operation conducted over a packed bed at 80°C, the following data are obtained. Determine the packing size and the column diameter.

Data: Surface tension of

Phenol at 80°C	= 37.3 dynes/cm
Lube oil at 80°C	= 26.8 dynes/cm
Extract	= 35.88 dynes/cm
Raffinate	= 29.14 dynes/cm

Density of

Phenol at 80°C	= 0.9692 gm/cm^3
Lube oil at 80°C	= 0.8161 gm/cm^3
Extract	= 0.8873 gm/cm^3
Raffinate	= 0.8043 gm/cm^3

Solution

d_{FC} is determined on the basis of properties of solvent and lube oil. From Eq. 20.170

$$\begin{aligned} d_{FC} &= 2.42\left[\frac{37.3-26.8}{(0.9692-0.8161)980}\right]^{0.5} \\ &= 0.6402 \end{aligned}$$

On the basis of the extract and raffinate phases:

$$\begin{aligned} d_{FC} &= 2.42\left[\frac{35.88-29.14}{(0.8873-0.8043)980}\right]^{0.5} \\ &= 0.6966 \end{aligned}$$

The higher size of packing is chosen. To avoid channeling, a minimum diameter of column required is eight times the critical packing size.

Example 20.13

Phenol extraction operation conducted at 100°C to extract aromatics from pressable waxy distillate, given the following data. Determine the number of stages used in the extractor.

Data

Feed, gm/min.	160
Solvent, gm/min	100
Extract, gm/min	110
Solvent free extract, gm/min	40.28
Aromatics (A_s) in the feed, %	21
Aromatics (A_s) in the extract, %	21.85

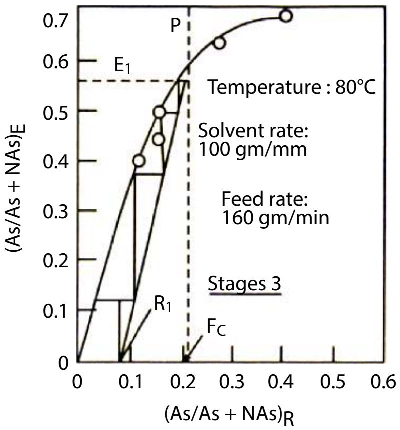


Figure 20.109 Determination of the number of extraction stages from Example 20.13.

Equilibrium data:

A_s in extract solvent free extract	A_s in extract solvent free raffinate
40.00	12.14
63.69	27.08
50.90	14.79
44.12	14.08
10.05	3.8

Solution

Total aromatics in the feed: $\frac{160 \times 21}{100} = 33.6$

Aromatics in the raffinate = 33.6 – 21.85 = 11.75

$(E_1) = \frac{A_s}{SFE} = \frac{21.85}{40.28} = 0.5426 \times 10^2$ (% A_s in the solvent free extract)

$(R_1) = \frac{A_s}{SFR} = \frac{11.75}{120} = 0.098 \times 10^2$ (% A_s in the solvent free raffinate)

The solvent free raffinate (SFR) = Feed-solvent free extract.

Figure 20.109 shows the equilibrium plot (x–y). In phenol extraction, the feed is introduced at the bottom of the column; the extract and feed are in contact (equilibrium) with each other, i.e., the feed concentration F_c = 0.21 and the extract concentration are in equilibrium. F_c and E_1 shall be fixing the point P. R_1 lies on x-axis, as the initial concentration of aromatics in solvent is zero. These two points are joined by a straight line, which forms the operating and equilibrium line.

Another method is by fixing the raffinate composition and drawing a straight line with slope equal to feed to solvent ratio until it touches the concentration of extract. From Figure 20.109, the number of stages is 3.

20.44 Phase Diagrams

Ternary liquid–liquid equilibrium data at fixed temperature and pressure are usually represented on an equilateral triangle or on a Janecke diagram (rectangular coordinates). These representations are generally associated respectively with a distribution diagram or a selectivity diagram. The distribution diagram gives the relationship between the solute content in the extract and in the raffinate in equilibrium. The distribution curve may intersect the first bisectrix and is referred to as a **solutrope**. The selectivity curve gives the relationship between the solute content in an extract and in a raffinate in equilibrium, excluding the solvent. When the selectivity curve intersects the first bisectrix, there is an **azeotrope**.

Analysis of liquid–liquid behavior and the design of an extraction cascade usually start with the phase equilibrium data. Table 20.63 shows data for the ternary system, acetic acid, water, and isopropyl ether. Each row in Table 20.63 corresponds to one tie line linking the organic (extract) and aqueous (raffinate) phase compositions at equilibrium. It is convenient to plot such data in an x–y diagram as shown in Figure 20.110. The solubility curve indicates the raffinate and extract compositions, and it defines the limits of the two-phase region. The extract and raffinate merge into a single phase at the plait point. Any point on Figure 20.110 defines one ternary composition. Points interior to the solubility diagram form two phases as illustrated in Figure 20.111, where a mixture with composition M in the diagram is shown to separate into two phases with extract composition E and raffinate composition R. Then inverse lever rule defines the relative amounts of the two phases. Given the line segment $\overline{EM}$ and $\overline{MR}$, relative amounts are

$$\frac{\overline{EM}}{\overline{MR}} = \frac{R}{E} = \frac{x_{wm} - x_{we}}{x_{wr} - x_{wm}} = \frac{x_{am} - x_{ae}}{x_{ar} - x_{am}} \tag{20.171}$$

where R and E are the masses of raffinate and extract, respectively, and x_{ij} are mass fractions with i = {w, a} referring to either with water or acetic acid mass fractions, and j = {e, m, r} referring to either point E, M, or R.

The point M is also known as the mix point and it defines the overall stoichiometric composition of the mixture (i.e., the composition if only a single phase was formed by mixing), while the extract and raffinate compositions E and R are found from constructing a line parallel to the nearest tie lines that also intersects the mix points and the solubility curve.

Table 20.63 Liquid–liquid equilibrium data obtained at 25°C and 1 atm.

LLE Data for Acetic Acid (A), Water (W), and Isopropyl Ether (E)					
Water-Rich Layer			**Ether-Rich Layer**		
Wt % A	**Wt % W**	**Wt % E**	**Wt % A**	**Wt % W**	**Wt % E**
1.41	97.1	1.49	0.37	0.73	98.9
2.89	95.5	1.61	0.79	0.81	98.4
6.42	91.7	1.88	1.93	0.97	97.1
13.30	84.4	2.3	4.82	1.88	93.3
25.50	71.1	3.4	11.4	3.9	84.7
36.70	58.9	4.4	21.6	6.9	71.5
45.30	45.1	9.6	31.1	10.8	58.1
46.40	37.1	16.5	36.2	15.1	48.7

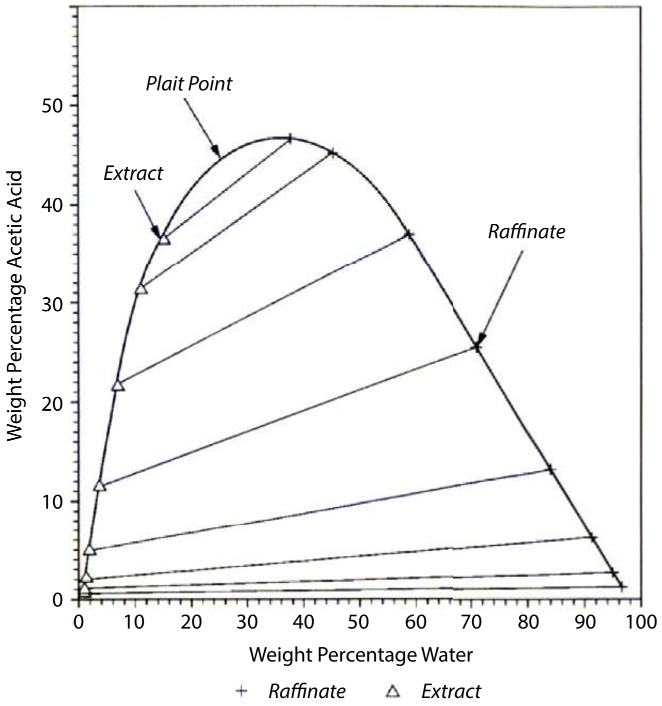


Figure 20.110 Liquid–liquid equilibrium data at 25°C for water, acetic acid and isopropyl ether, showing raffinate and extract compositions at several tie lines and the plait point. At the plait point, both phases dissolve into a single phase with the same composition. The weight percentage of isopropyl ether equals 100 less the weight percentage of water less the weight percentage of acetic acid.

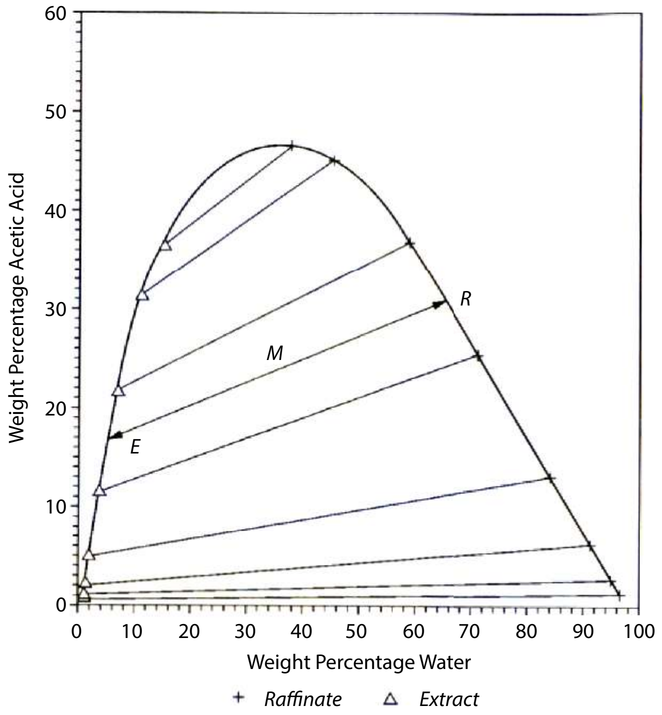


Figure 20.111 Liquid–liquid data for the system water, acetic acid and isopropyl ether. Since mixture M falls inside the solubility curve, it separates into extract and raffinate phases with compositions E and R respectively. The equilibrium data are at 25°C and 1 atm.

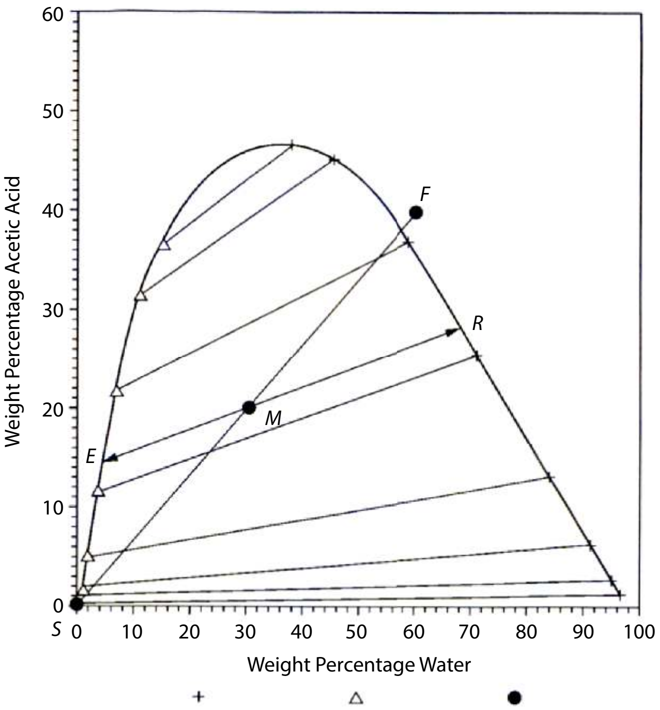


Figure 20.112 Liquid – liquid data at 25°C for the system water, acetic acid, and isopropyl ether. Mixing liquids with compositions F and S in equal proportions by mass results in mix point M, which separates into phases with compositions E and R. The equilibrium data are at 25°C and 1 atm.

The compositions of feeds and solvent mixtures may lie outside the solubility curve; however, if upon mixing, the overall stoichiometry is such that the M point falls inside the solubility curve, then two liquid phases will form. Mixtures outside of the solubility curve are only a single phase. Therefore, Figure 20.111 indicates that the addition of acetic acid to water increases the solubility of isopropyl ether in both the extract and raffinate phases. Aqueous mixtures containing 50 mass % of more of acetic acid are miscible with isopropyl ether in all proportions.

Consider mixing equal amounts of pure isopropyl ether with an aqueous feed that is 40 mass % acetic acid in water. The mix point M then lies at a point equidistant between the feed composition point F and the solvent composition point S. The resulting two phases have the compositions E and R as shown in Figure 20.112.

20.45 Countercurrent Extractors

Cascades of countercurrent extractors as shown in Figure 20.113 give the highest recover of solute while requiring the least amount of solvent [114]. The alternatives, cocurrent, and crosscurrent extractions, are less used. Steady-state material balances written around the jth stage in Figure 20.113 result in

$$R_{j-1}\,x_{j-1} + E_{j+1}\,y_{j+1} = R_j x_j + E_j y_j \tag{20.172}$$

For a cascade consisting of n stages, R_1 is the feed and E_{n+1} is the fresh solvent added to the cascade, or column at the opposite end. The mix point, M, can then be defined in terms of

$$M = E_0 + R_{n+1} = E_n + R_1 \tag{20.173}$$

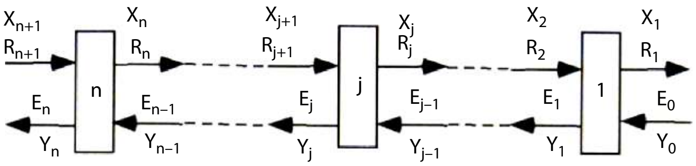


Figure 20.113 Stages in a countercurrent cascade. R_{n+1} is the aqueous feed. E_0 is the fresh solvent.

which can be arranged to give the difference point D,

$$D = E_0 - R_1 = E_n - R_{n+1} \tag{20.174}$$

As shown in Figure 20.114, the line $E_0 - R_1$ intersects the line $E_n - R_{n+1}$ at the D-point. Graphically, all lines that intersect the D-point and a raffinate composition on the solubility curve between R_1 and R_{n+1}, say R_{j+1}, also intersect the extract composition, E_j, on the solvent-rich side of the solubility curve. Each line that intersects the D-point and a raffinate composition, R_{j+1}, also intersects the extract, E_j, which is passing that raffinate on the jth stage. If the extract E_j and raffinate R_j leaving the j^{th} stage are assumed to be at equilibrium, then a graphical method can be used to step off theoretical stages as shown in Figure 20.115. Graphical methods are suitable when two liquid phases are partly miscible as for the water, acetic acid, and isopropyl ether system. If the two liquid phases are nearly immiscible, and their solubility is not affected by the solute concentration, then the Kremser equation may be applied.

20.45.1 Kremser Equation

Liquid–liquid systems that obey Nernst's law can be simple to analyze since the equilibrium relationship is linear in composition. If the two liquid phases are nearly immiscible, and if the solute concentrations are low in either phase,

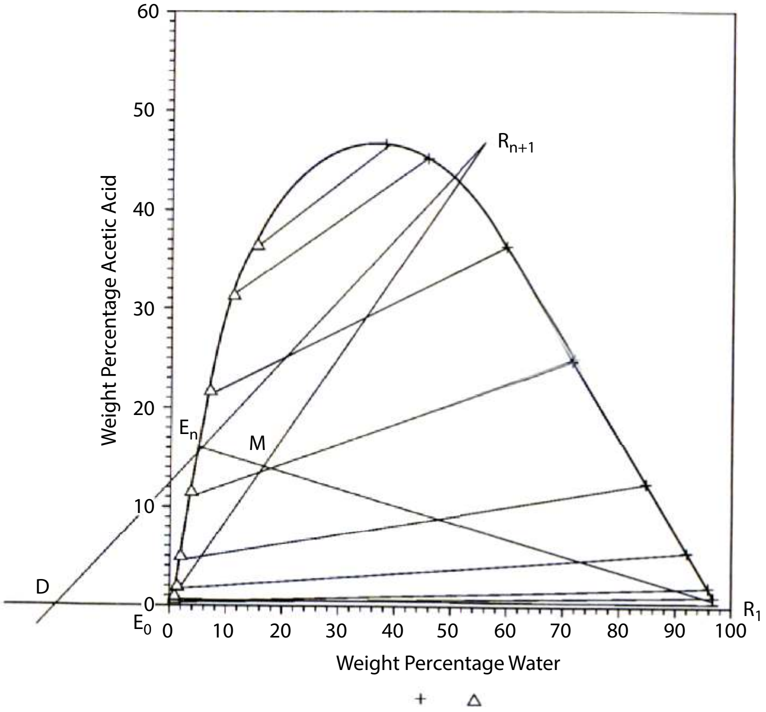


Figure 20.114 Liquid–liquid data at 25°C for the system water, acetic acid and isopropyl ether. Mixing liquids with compositions R_{n+1} and E_0 results in mix point M. Specification of R_1 yields E_n since the same line also intersects M. The equilibrium data are at 25°C and 1 atm.

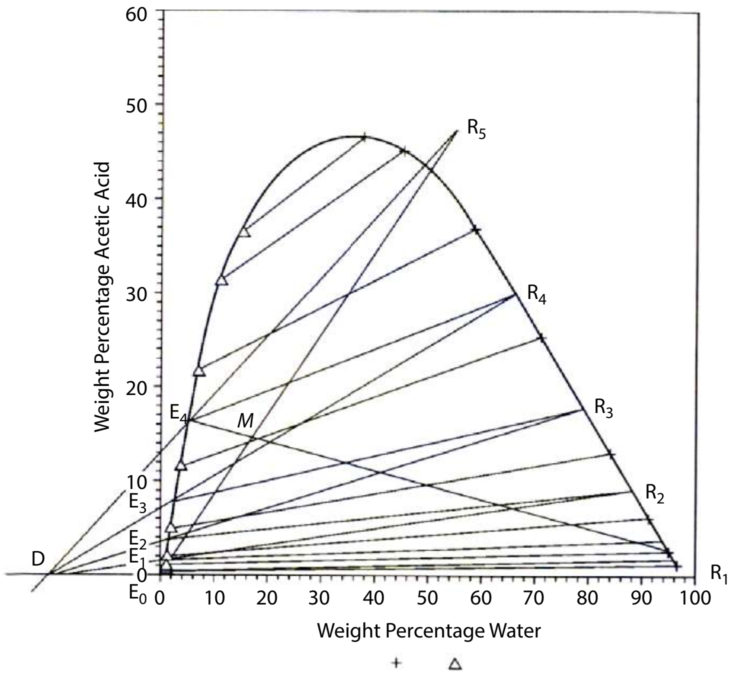


Figure 20.115 Liquid–liquid data at 25°C for the system water, acetic acid and isopropyl ether. Theoretical stages are stepped off by either drawing lines through D to obtain an extract composition or interpolating between nearby tie lines to estimate the extract composition on the next stage. The equilibrium data are at 25°C and 1 atm.

then the operating or material balance line is also nearly linear. The material balance around the j^{th} stage and the raffinate end of the cascade in Figure 20.113 gives the operating line

$$y_i = \frac{R_{j+1}}{E_j} x_{j+1} + \frac{E_0 y_0 - R_1 x_1}{E_j} \tag{20.175}$$

where Ej and Rj are the extract and raffinate flow rates leaving the jth stage, usually assumed to be in thermodynamic equilibrium. If $E_0 \approx E_1 \approx$ ……..E_j and $R_1 \approx R_2$ ……..R_{j+1}, then the subscripts on E and R may be dropped, and the ratio R/E is nearly constant. In this case, the operating line Eq. 20.175 is linear and $(Ey_0 - Rx_1)/E$ is constant. Under these conditions, the extraction factor $E_e = K_{DI}\,E/R$ or $E_e = D_i\,E/R$ is also constant*, and the Kremser [115] group method applies:

$$\phi_E = \frac{E_e - 1}{E_e^{N+1} - 1} \tag{20.176}$$

where ϕ_E is the solute fraction in the entering feed that is not extracted, N is the required number of theoretical (or equilibrium) stages to extract $1 - \phi_E$ fraction of the feed, and E_e is the extraction factor, which is constant. Figure 20.116 shows the Kremser plot.

---

* E_e is dimensionless. If K_{Di} is used, then E/R is a molar ratio. If D_i is used, then E/R is a mass ratio.

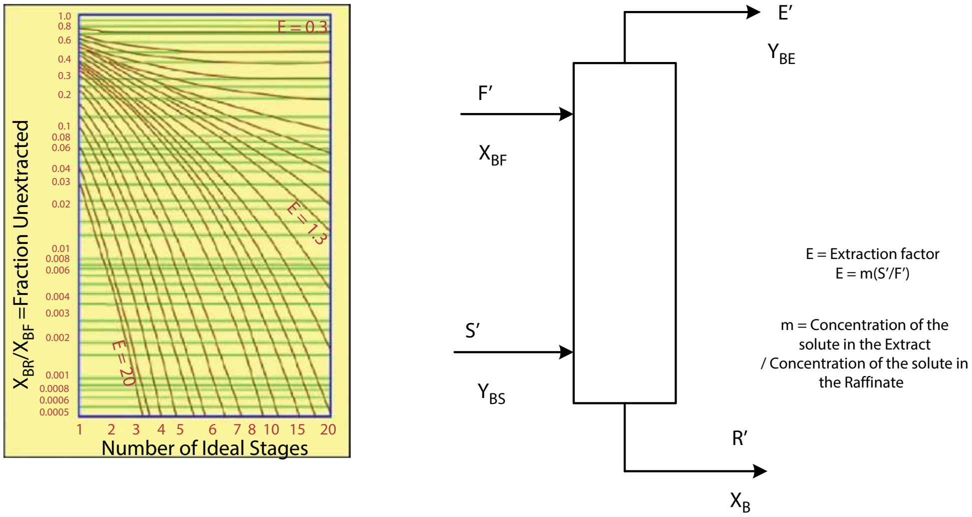


Figure 20.116 The Kremser plot with Fraction unreacted vs. number of stages (source: Koch Modular Process System (KMPS) LLC).

Rearranging Eq. 20.176 and solving for N gives

$$N = \frac{\ln\left[\dfrac{(E_e - 1)}{\phi_E} + 1\right]}{\ln E_e} - 1 \tag{20.177}$$

The Kremser equation is defined by

$$N = \frac{\log\left[\left(\dfrac{x_f - \dfrac{y_s}{m}}{x_n - \dfrac{y_s}{m}}\right)\left(1 - \dfrac{1}{E}\right) + \dfrac{1}{E}\right]}{\log E} \tag{20.178}$$

where

N = number of theoretical stages required
x_f = concentration of solute in feed on solute free basis
x_n = concentration of solute in raffinate on solute free basis
y_s = concentration of solute in solvent on solute free basis
m = distribution coefficient
E = Extraction factor (m) (S/F)
S = Solvent rate
F = Feed rate

Example 20.14

A countercurrent LLE cascade is required to remove 99% of solute species A from an aqueous phase. The aqueous feed rate is 100 kg/h, and the feed contains 5 mass % species A. The distribution coefficient on a mass ratio basis is nearly constant, and D_i equals 1.4 kg water/kg solvent. Determine the number of theoretical stages required if pure solvent can be provided to cascade at a rate of 90 kg/h.

Solution

$E_e = D_i\,E/R = 1.4(90)/100 = 1.26$ and the fraction not extracted is $\phi_E = 0.01$ From Eq. 20.177, N gives

$$N = \frac{\ln\left[\frac{(1.26-1)}{0.01}+1\right]}{\ln 1.26} - 1$$

N = 13.24 theoretical stages.

20.46 Extraction Equipment

For the purpose of an extraction process, the feed and the solvent should be in close contact and then be separated again (usually by gravity). This can be done either in a continuous or in a batch equipment.

Mixer-Settler

The mixer-settler is the simplest form of an extractor. If the extraction is carried out in batches, it may be conducted in a single vessel. However, it can be conducted in a continuous process as shown in Figure 20.117. This form of extractor is also referred to as mixer-settler. In the mixer, there is a stirrer for the thorough mixing of the phases. Then the mixture is gradually passed to the separator, the mixing being continuous. The combination of the two processes is referred to as one stage. The extract is removed with the aid of compressed air. Figure 20.118 shows some arrangement mixer-settler types in solvent extraction processes.

A mixer-settler has the following:

Advantages:

1. The stirrer allows a good control of the degree of distribution, whereas control with extraction columns depends heavily on the liquid streams.
2. Finely distributed solid particles are less of a nuisance than with other forms of extraction.

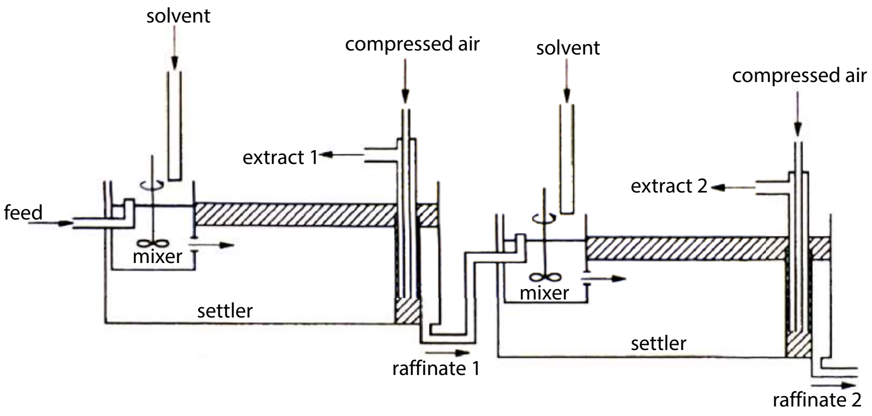


Figure 20.117 Mixer-settler (two-stage continuous extraction).

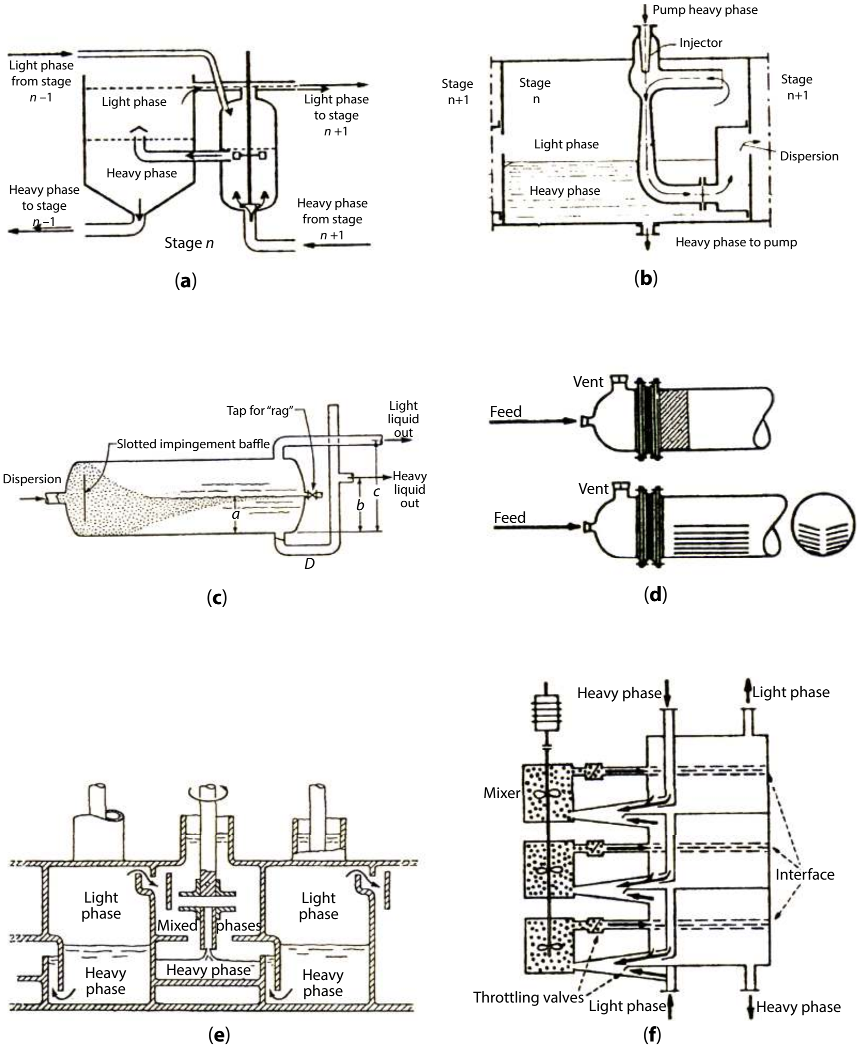


Figure 20.118 Some types and arrangements of mixers and settlers. (a) Kermira mixer-settler (Mattila, Proc. Solvent Extraction Conference, ISEC 74, Inst. Chem. Eng., London 1974); (b) Injection mixer and settler (Ziolkowski, 1961). (c) Gravity settler; "rag" is foreign material that collects at the interface. (d) Provisions for improving rate of settling: (top) with packing or wire mesh; (bottom) with a nest of plates. (e) Compact arrangement of pump mixers settlers [Coplan *et al*. Chem. Eng. Prog., 50, 403 (1954)]. (f) Vertical arrangement of a battery of settlers and external mixers (Lurgi Gesellschaften).

3. When the process is interrupted, the liquids remain in the stage where they are.
4. The number of stages can be increased very easily.
5. High stage efficiencies can be reached and scaling up of the process is easy.

Disadvantages:

1. It requires a lot of floor space.
2. Between the stages, the liquid must be pumped frequently.

Columns

Liquid–liquid extraction can also be conducted in columns. They all work according to the principle of countercurrent extraction and are mounted vertically. Like other physical separation methods, the extraction in columns is based on the difference in density between the two liquids and on gravity. The liquid having the higher density is therefore fed to the top of the column and the liquid with the lower density to the bottom. They flow in opposite directions, to be carried off at the other end of the column. To obtain a thorough mixing of the liquids, various techniques are used. For instance, empty columns are packed with packing material or provided with sieve plates, rotating discs, or bubble trays. Figure 20.119 shows column types without agitation.

Sieve tray columns are popular in the chemical and petrochemical industries. The trays minimize axial mixing, which results in good scale-up from laboratory data. The dispersed phase drops re-form at each perforation, rise (or fall) near their terminal velocity, and then coalesce underneath (or above) the tray, as illustrated in Figure 20.119d. The coalesced layer is important to prevent axial mixing of the continuous phase and to allow re-formation of the drops, which increases mass transfer. The continuous phase passes through the downcomer (upcomer) and across the sieve tray. The height of the coalesced layer depends on the combined pressure drop of the dispersed phase through the perforations and the continuous phase through the downcomer (upcomer) [116].

In commercial sieve tray extractors, the height of the coalesced layer should be designed for 1–2 in. In general, the pressure drop should be balanced between the downcomer and the orifice. Columns with sieve plates as shown in Figure 20.120 or packing materials are identical with the equipment that is used in absorption and rectification. The bubble cap tray columns used in rectification are hardly ever or never applied in extraction because of the high-pressure loss that liquids experience on account of the bubble caps.

Rotating Disc Contactor

In the rotating disk contactor (RDC) extractor, the impellers are flat disks; the mixing zones are separated by partial diametral baffles referred to as stators, but the distinct settling zones are not provided. The RDC is one of the most

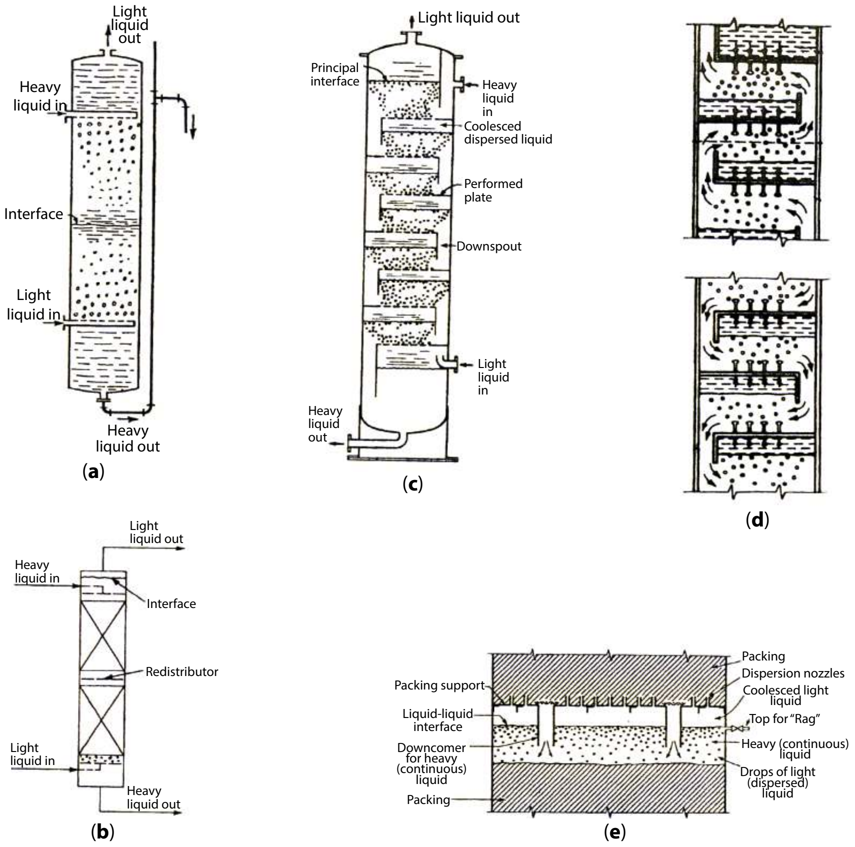


Figure 20.119 Tower extractors without agitation. (a) Spray tower with both phases dispersed. (b) Two-section packed tower with light phase dispersed. (c) Sieve tray tower with light phase dispersed. (d) Sieve tray construction for light phase dispersed (left) and heavy phase dispersed (right). (e) Redistributor for packed tower with light phase dispersed (Treybal, 1963).

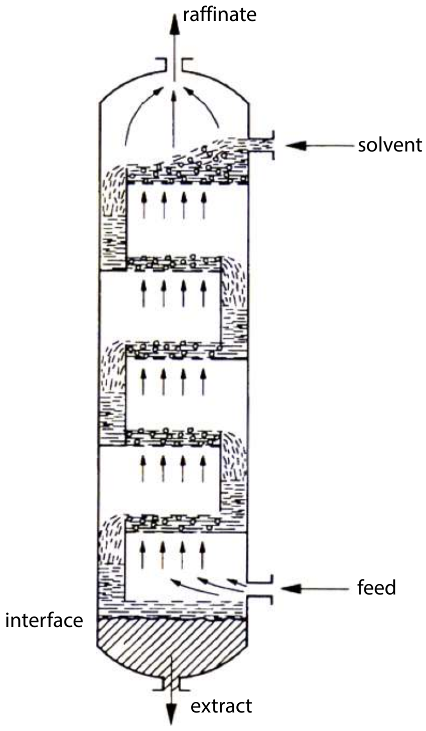


Figure 20.120 Sieve plate column.

widely employed of agitated extractors because of its geometrical simplicity and effectiveness. Situations where it may not be suitable are when only a few stages are required, and as such the mixer-settlers will be satisfactory and cheaper or when their large holdup and long residence times may be harmful to unstable substances. Furthermore, for systems with low interfacial tensions and low-density differences and as such, stable emulsions may be formed by the intense agitation.

Contact between extracts and raffinate can be promoted by stirring the contents of the columns or agitating them in some other way as shown in Figure 20.121. In such a column, a number of compartments are formed by horizontal solid plates with a circular opening in the middle. In the heart of the column, there is a rotating shaft, on which discs are mounted. The diameter of the disks is smaller than that of the openings in the horizontal plates.

The supply of both solvent and feed is done tangentially, in the same direction as the direction of rotation of the shaft. The liquid having the higher density is introduced near the top and the liquid with the lower density near the bottom.

The centrifugal force exerted by the stirrer and the way in which the two liquids are introduced ensure a good contact between them. The separation into liquid layers can (depending on the operating system) take place either at the top or at the bottom of the column. The rotary speed of the shaft can usually be varied. The degree of contact between the two liquids can be adjusted to some extent.

Advantages:

1. The very high extraction efficiency.
2. RDCs are suitable for very high throughputs.
3. They can be made very high (up to a few dozen meters), so that extractions with a large number of theoretical stages are within reach.

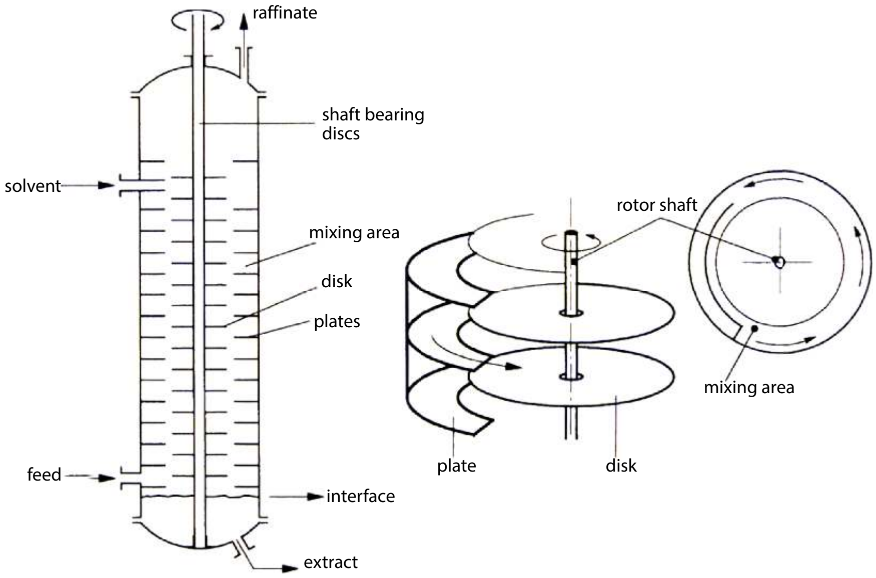


Figure 20.121 Rotating disc contactor.

Disadvantages:

1. The apparatus is expensive.
2. Cleaning is difficult; the apparatus is easily fouled.

Figure 20.122 shows a flow chart of major types of extraction equipment, and Figure 20.123 illustrates towers with rotary agitators. Figure 20.124 depicts towers with reciprocating trays or with pulsing action, and Figure 20.125 illustrates a block diagram of successive extraction applications, respectively. Figure 20.126 provides a comparison of commercial extractors based on efficiency/stages per meter vs. capacity.

According to Figure 20.126, the RDC is intermediate in stage efficiency and throughput. The value of HETS = 0.3 m, and this compares roughly with the HTU = 0.4 or 0.75 depending on which phase is dispersed of the pilot plant data. The design procedure used by Kosters, of Shell Oil Co., who developed this equipment requires pilot plant measurements on the particular system of HTU and slip velocity as functions of power input. The procedure for scale-up and the results of a typical design worked out by Kosters (in Lo *et al*. 1983, pp 391–405) are summarized in Example 20.13. Scale-up by this method is reliable from 64 mm diameter to 4–4.5 m diameter. The data of Figure 20.128 are used in the study.

Example 20.15: Design of a Rotating Disk Contactor (RDC)

A hydrocarbon mixture containing 10% aromatics and at the rate of 55.5 m3/h is to be treated with a solvent at the rate of 173.6 m^3/h. Ten stages are needed for the extraction. Pilot plant data are available for the HTU and the slip velocity; they are shown on the graphs for solvent either continuous or dispersed. The procedure of Table 20.64 was applied by Kosters (in Lo *et al.* pp 391–405, 1983) with the following results.

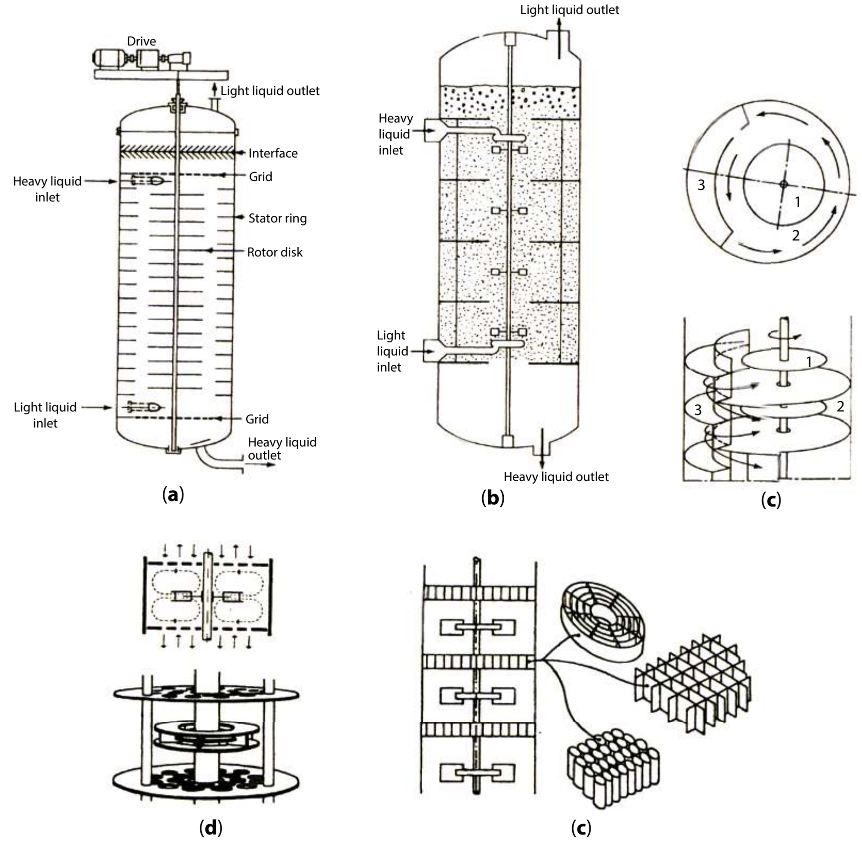


Figure 20.122 Tower extractors with rotary agitators. (a) RDC (rotating disc contactor) extraction tower (Escher B.V. Holland). (b) Oldshue–Rushton extractor with turbine impellers and stator rings (Mixing Equip. Co.). (c) ARD (asymmetric rotating disk) extractor: (1) rotating disk rotor; (2) mixing zone; (3) settling zone (Luwa A.G.). (d) Kuhni extractor, employing turbine impellers and perforated partitions (Kuhni Ltd.). (e) EC (enhanced coalescing) extractor [Fischer *et al.*, Chem. Ing. Tech. 228 (Mar. 1983)].

	Solvent Continuous	**Solvent Dispersed**
Vessel dia. (m)	2.1	1.7
Stator dia. (m)	1.47	1.19
Rotor dia. (m)	1.26	1.02
HTU (m)	0.41	0.75
(HTU)off (m)	0.663	1.107
Number of compartments	40	81
Compartment height (m)	0.20	0.17
Total height (m)	10.4	15.7
Rotor speed (rpm)	15-60	15-70
Power (theoretical kW)	4.6	2.8

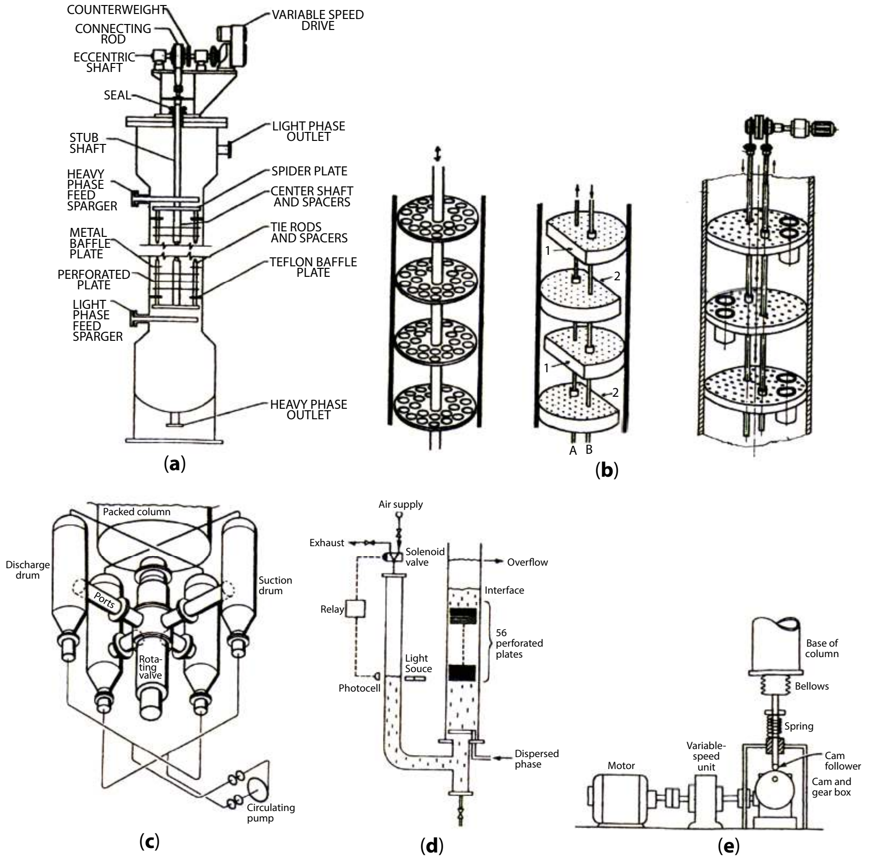


Figure 20.123 Towers with reciprocating trays or with pulsing action. (a) Assembly of a 36 in. Karr reciprocating tray column (Chem. Pro. Co.). (b) Sieve trays used in reciprocating trays columns; (left) large opening trays for the Karr column; (middle) countermotion trays with cutouts; (right) countermotion trays with downpipes for heavy phase. (c) Rotary valve pulsator, consisting of a variable speed pump and a rotary valve that alternately links the column with pairs of suction and discharge vessels. (d) Sieve tray tower with a pneumatic pulser [Proc. Int. Solv. Extr. Conf. 2, 1571 (1974)]. (e) A pulser with a cam-operated bellows.

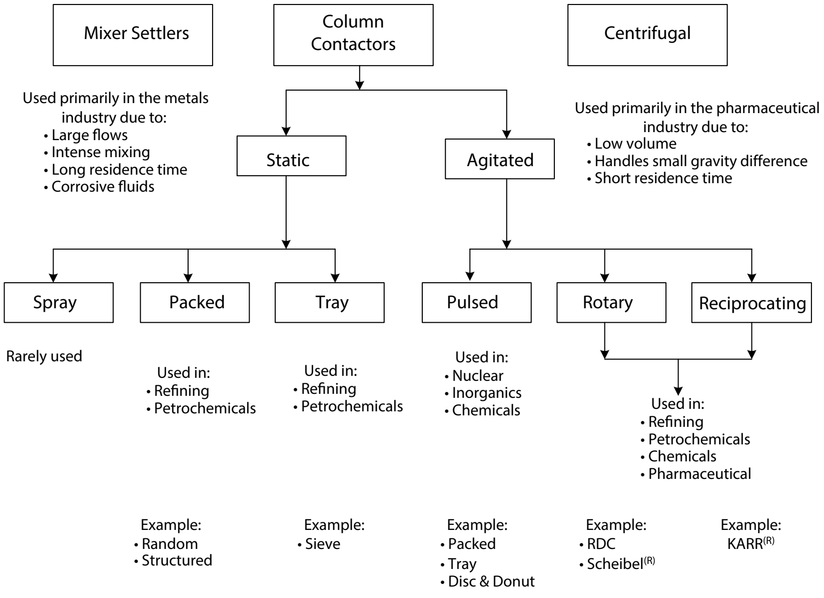
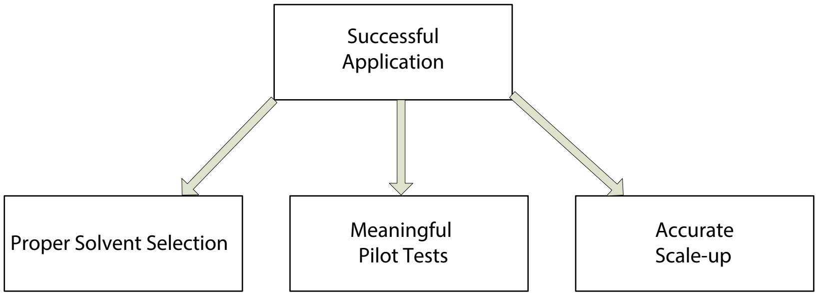


Figure 20.124 Major types of extraction equipment.



Selection based on :
- Sound thermodynamic principles
- Sound economic principles
- Availability
- Recoverability
- Sound environmental principles
- Toxicity
- Safety

Testing based on:
- Actual feed stocks
- Full process including solvent recovery
- Wide range of operating conditions

Scale-up based on:
- Proven techniques
- Proper safety factors

Figure 20.125 A block diagram of successive extraction applications.

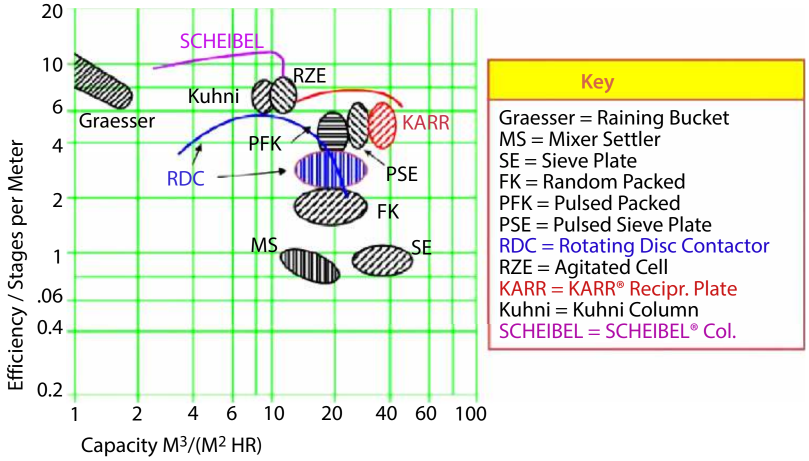


Figure 20.126 Efficiency and capacity range of small diameter extractors, 50–150 mm dia. Acetone extracted from water with toluene as the dispersed phase (source: Koch Modular Process System (KMPS) LLC).

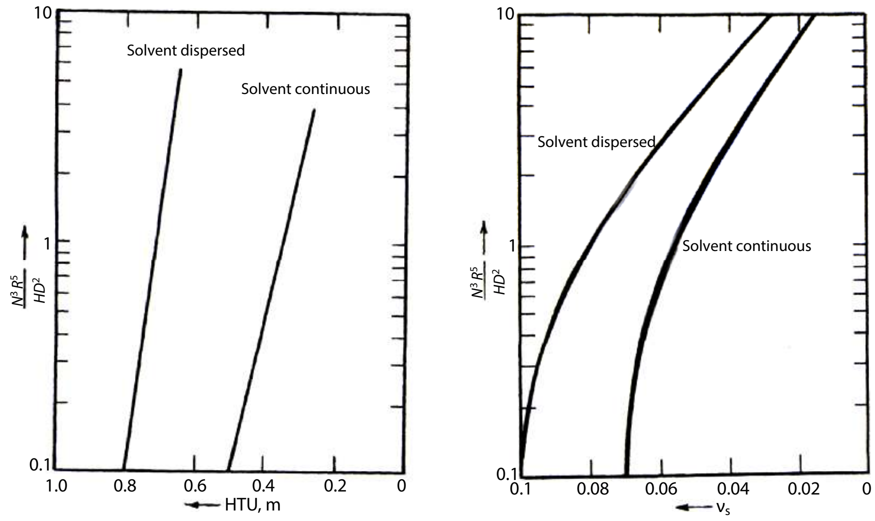


Figure 20.127 Hold up at flooding, power input, and slip velocity in an RDC (Kosters, in Lo., Baird and Hanson, 1983).

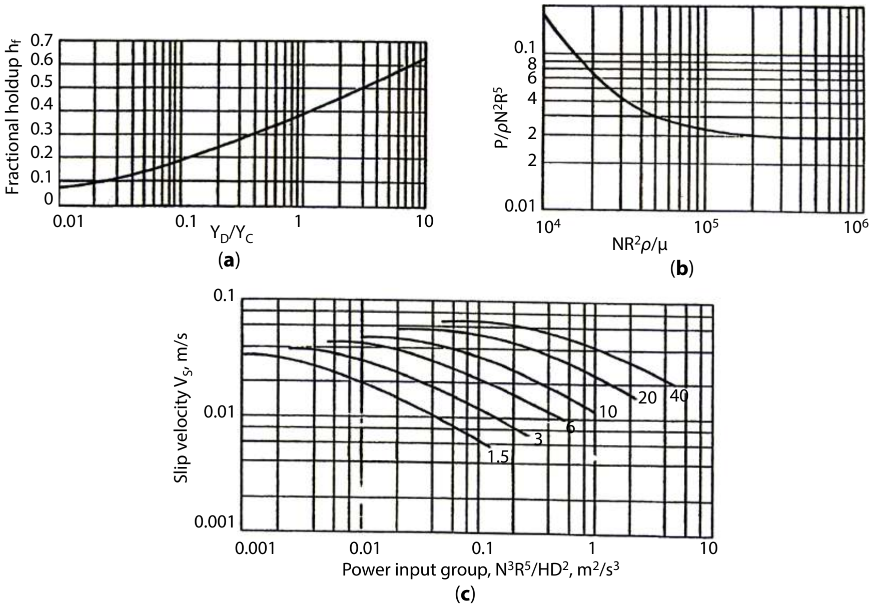


Figure 20.128 Fractional holdup at flooding, hf, as a function of flow ratio of the phases. (b) Power input to one rotor as a function of rotation speed N and radius R. (c) Slip velocity versus power input group of density difference of 0.15 g/ml at the indicated surface tensions (dyne/cm).

Table 20.64 Formulas for sizing an RDC.

1. Stator opening diameter, S = 0.7D, where D is vessel diameter	8. Holdup h at an opening velocity V_C say 70–80% of flooding, $V_{Cf} = \dfrac{V_s \exp(-h)}{V_D/V_C\, h + 1/(1-h)}$ Solve by trial for h when other quantities are specified.
2. Rotor diameter, R = 0.6D	9. Effective height of a transfer unit $(HTU)_{off} = (HTU)_{pilot\ plant}$ + HDU in terms of a value obtained in a pilot plant and a calculated height of a diffusion unit (HDU)
3. Height to diameter ratio of a compartment. D(m) 0.5–1.0 1.0–1.5 1.5–2.5 2.5 H/D 0.15 0.12 0.1 0.08–0.1	10. Height of a diffusion unit, $HDU = H(1/Pe_C + 1/Pe_D)$
4. Power input, Figure 20.127	11. Factors E_C and E_D for evaluating the Peclet numbers, $E_C = 0.5V_CH + 0.012RNH(S/D)^2$ $E_D = E_C[4.2(10^5)(V_D/h)^{3.3}/D^2]$ when the correction in brackets is less than unity, make $E_D = E_C$
5. Fractional holdup at flooding, h_f, from Figure 20.128a.	12. The Peclet numbers are $1/Pe_C = E_C(1-h)/HV_C$ $1/Pe_D = E_D h/HV_D$

(*Continued*)

Table 20.64 Formulas for sizing an RDC. *(Continued)*

6. Slip velocity V_s preferably is obtained experimentally, but is given approximately by Figure 20.128c	13. Final expression for height of a diffusion unit is $HDU = E_C(1-h)/V_C + E_d h/V_D$
7. Superficial velocity of the continuous phase at flooding, $V_{Cf} = \dfrac{V_s \exp(-h_f)}{V_D/V_C\,h_f + 1/(1-h_f)}$ where V_C and V_D are the superficial velocities of the continuous and dispersed phases.	

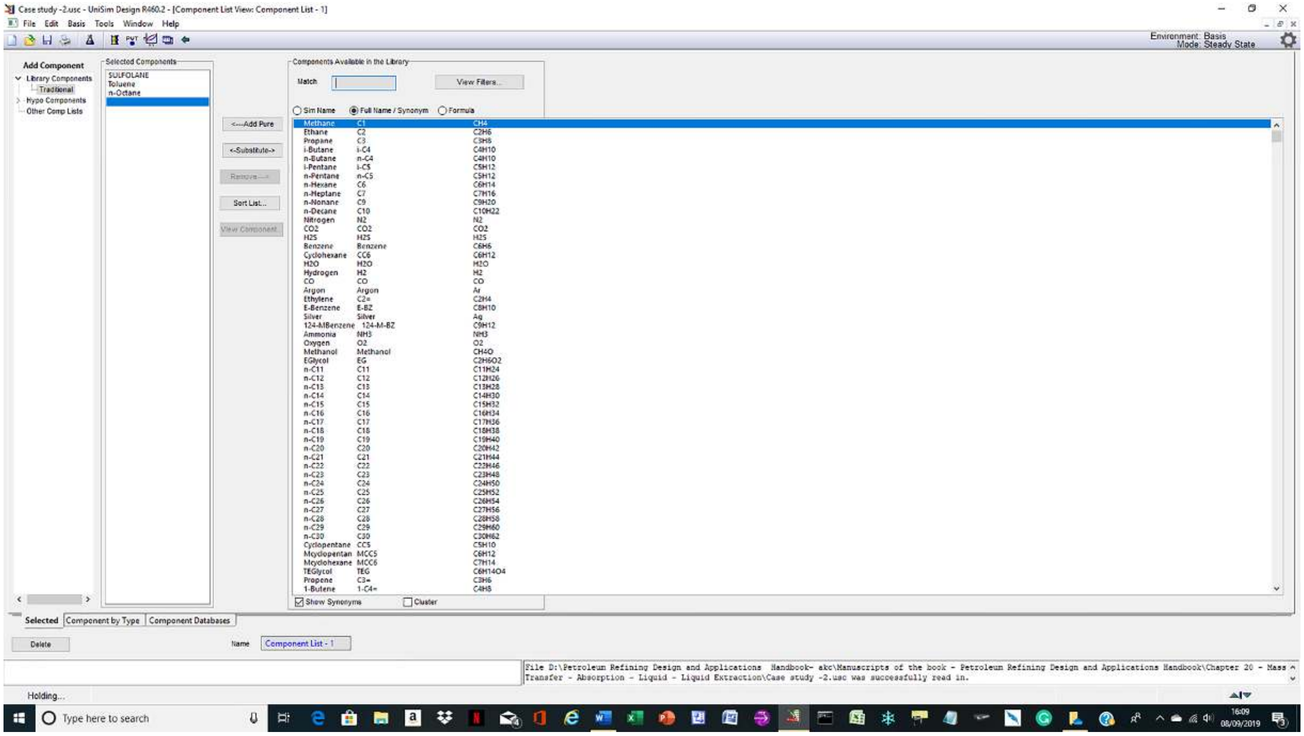
(W.C.G. Kosters, Shell Oil Co.)

Case Study 2

An extractor is to be designed such that toluene will be extracted from a feed mixture of 0.5 mole fraction of toluene and 0.95 mol fraction of n-octane and at a feed rate of 100 kg/h. The temperature and pressure are 25°C and 150 kPa respectively. Sulfolane will be used to extract the toluene at a composition of 1.0 mole fraction and at a feed rate of 100 kg mole/h. The column has 15 plates and the pressure at the bottom is 250 kPa. Determine the composition of the extract and raffinate phases.

Solution

UniSim Design R460.2 (Case study 2.usc) software was used to extract the solute, toluene in the feed of toluene and n-octane, and the solvent Sulfolane with the solvent at the top of the column and the feed being fed to the bottom of the column having 15 stages, operating at a temperature of 25°C, and a ΔP = 100 kPa.



Case study -2.usc - UniSim Design R460.2 - [Component List View: Component List - 1]

Selected Components
SULFOLANE
Toluene
n-Octane

Components Available in the Library

Match

○ Sim Name ● Full Name / Synonym ○ Formula

Methane	C1	CH4
Ethane	C2	C2H6
Propane	C3	C3H8
i-Butane	i-C4	C4H10
n-Butane	n-C4	C4H10
i-Pentane	i-C5	C5H12
n-Pentane	n-C5	C5H12
n-Hexane	C6	C6H14
n-Heptane	C7	C7H16
n-Nonane	C9	C9H20
n-Decane	C10	C10H22
Nitrogen	N2	N2
CO2	CO2	CO2
H2S	H2S	H2S
Benzene	Benzene	C6H6
Cyclohexane	CC6	C6H12
H2O	H2O	H2O
Hydrogen	H2	H2
CO	CO	CO
Argon	Argon	Ar
Ethylene	C2=	C2H4
E-Benzene	E-BZ	C8H10
Silver	Silver	Ag
124-MBenzene	124-M-BZ	C9H12
Ammonia	NH3	NH3
Oxygen	O2	O2
Methanol	Methanol	CH4O
EGlycol	EG	C2H6O2
n-C11	C11	C11H24
n-C12	C12	C12H26
n-C13	C13	C13H28
n-C14	C14	C14H30
n-C15	C15	C15H32
n-C16	C16	C16H34
n-C17	C17	C17H36
n-C18	C18	C18H38
n-C19	C19	C19H40
n-C20	C20	C20H42
n-C21	C21	C21H44
n-C22	C22	C22H46
n-C23	C23	C23H48
n-C24	C24	C24H50
n-C25	C25	C25H52
n-C26	C26	C26H54
n-C27	C27	C27H56
n-C28	C28	C28H58
n-C29	C29	C29H60
n-C30	C30	C30H62
Cyclopentane	CC5	C5H10
Mcyclopentan	MCC5	C6H12
Mcyclohexane	MCC6	C7H14
TEGlycol	TEG	C6H14O4
Propene	C3=	C3H6
1-Butene	1-C4=	C4H8

☑ Show Synonyms ☐ Cluster

Name: Component List - 1

```
File D:\Petroleum Refining Design and Applications  Handbook- akc\Manuscripts of the book - Petroleum Refining Design and Applications Handbook\Chapter 20 - Mass
Transfer - Absorption - Liquid - Liquid Extraction\Case study -2.usc was successfully read in.
```

Figure 20.129 A snap-shot of Selected page of the Component List View page in Component List Window showing the selected components (courtesy of Honeywell Process Solution, UniSim Design R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).

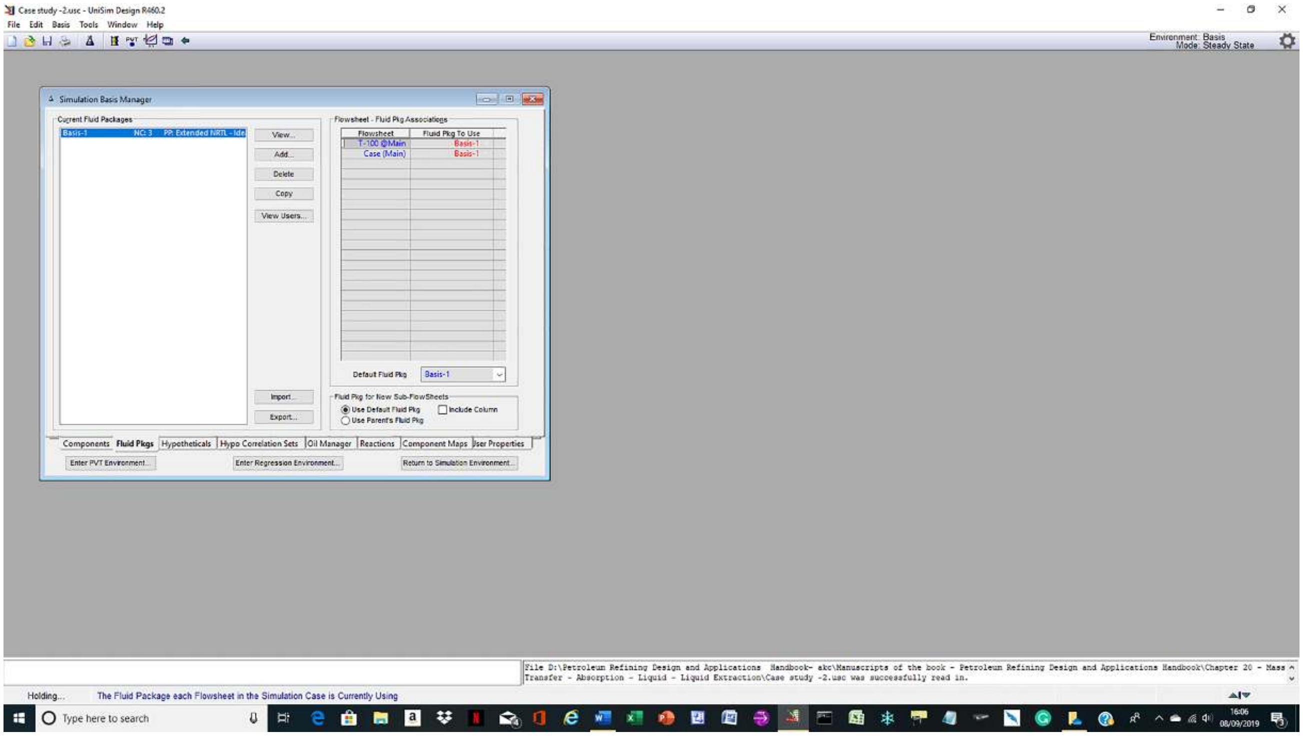
In liquid–liquid extraction, there must be two separate liquid phases; however, the thermodynamics predicts only one liquid phase when mixing the Feed and Sulfolane streams in a mixer and viewing the resulting outlet. Figures 20.129 and 20.130 show snapshots of the **Selected** page of the Component List View: Component List-1 Window showing the selected components and the Fluid Pkgs page of the Simulation Basis Manager showing the Extended NRTL—Ideal in Simulation Basis Manager Window, respectively.

Click on the **Home View** icon as shown in Figure 20.131 to display the Fluid page on the Basis Manager Window. Clicking on **View** button to display the Activity Model Interaction Parameters as shown in Figure 20.132. There are no values Binary Interaction Parameters (BIPs) for Sulfolane–n-Octane component pair. These can be estimated by using UNIFAC by selecting **UNIFAC LLE** and Press the "**Unknowns Only**" button to generate the Binary Coefficients as shown in Figure 20.132. The Binary Interaction Parameters (BIPs) are tuning parameters for the fluid pkg. For activity coefficients models like NRTL, it is required that the user supplies BIPs. BIPs essentially change the results of the phase equilibrium, and as such, it is essential that the phase equilibrium results are validated by experimental data before being used in any simulation. Note that if the phase equilibrium is wrong (i.e., does not match realty), then the simulation results would be invalid.

A mixer icon is used in the simulation, as the aim is to establish what happens at the outlet when the feed stream is in contact with the solvent. From this, UniSim Design simulation predicts only a single liquid phase; thus a liquid–liquid extractor would not work. UniSim Design has a feature called multiple streams, which can be enabled in **Preferences (Tools/Preferences, Simulation tab, Options page)** as illustrated in Figure 20.133. Two liquid phases are determined when the Feed and Sulfolane streams are mixed as shown in Figure 20.134.

The column convergence depends on the number of stages and the feed conditions as shown in Figure 20.135.

Note: The simulation converges when the solvent Sulfolane is fed at the top of the column and the Feed is at the bottom; otherwise, it will not solve. Figure 20.135 shows the snapshot of process flow diagram of the liquid–liquid extraction simulation. Figures 20.136 and 20.137 show temperature and the pressure versus tray position from the top of the column, respectively. The results from the simulation (Case study 2.usc) are


Case study -2.usc - UniSim Design R460.2

Environment: Basis
Mode: Steady State

Simulation Basis Manager

Current Fluid Packages

Basis-1 NC: 3 PP: Extended NRTL - Ide

Flowsheet - Fluid Pkg Associations

Flowsheet	Fluid Pkg To Use
T-100 @Main	Basis-1
Case (Main)	Basis-1

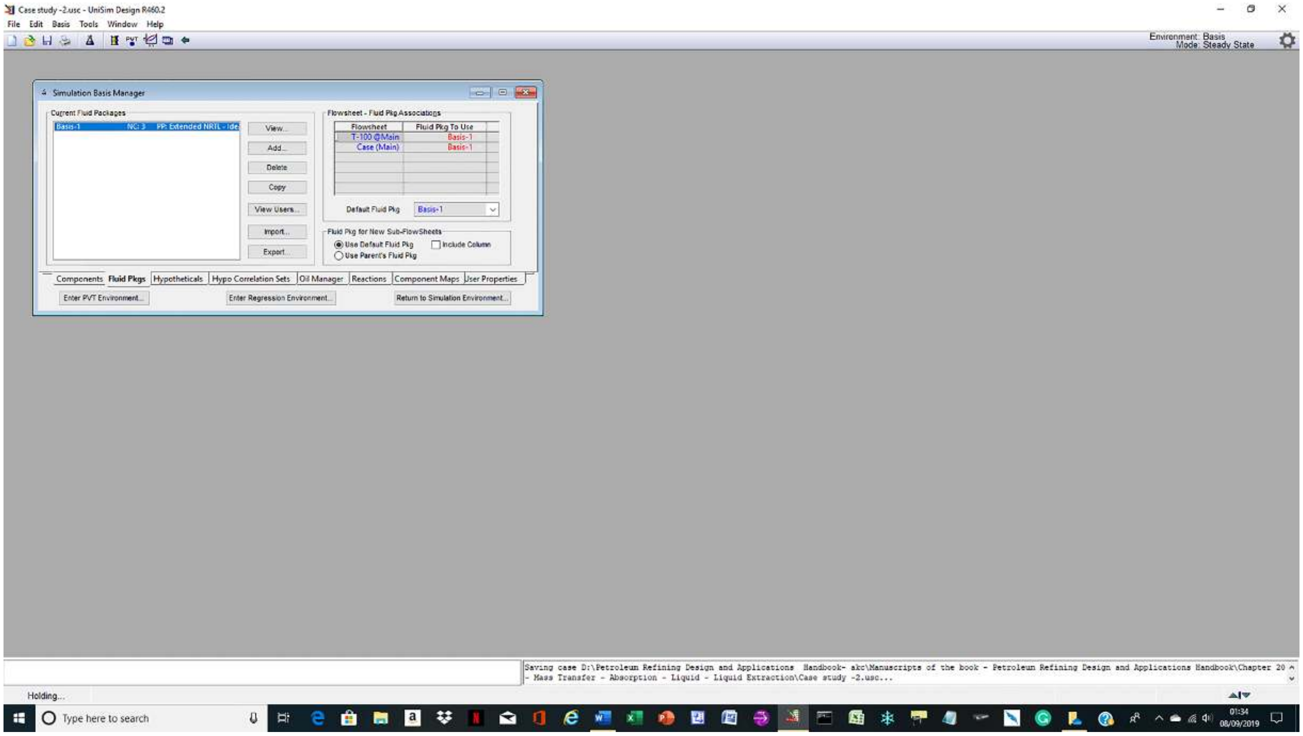
Default Fluid Pkg: Basis-1

Fluid Pkg for New Sub-FlowSheets
● Use Default Fluid Pkg ☐ Include Column
○ Use Parent's Fluid Pkg

```
File D:\Petroleum Refining Design and Applications  Handbook- akc\Manuscripts of the book - Petroleum Refining Design and Applications Handbook\Chapter 20 - Mass
Transfer - Absorption - Liquid - Liquid Extraction\Case study -2.usc was successfully read in.
```

Holding... The Fluid Package each Flowsheet in the Simulation Case is Currently Using

Figure 20.130 A snap-shot of Fluid Pkgs page of the Simulation Basis Manager Window showing the Extended NRTL – Ideal in Simulation Basis Manager Window (courtesy of Honeywell Process Solution, UniSim Design R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



Current Fluid Packages

Basis-1 NC: 3 PP: Extended NRTL - Ide

Flowsheet - Fluid Pkg Associations

Flowsheet	Fluid Pkg To Use
T-100 @Main	Basis-1
Case (Main)	Basis-1

Default Fluid Pkg: Basis-1

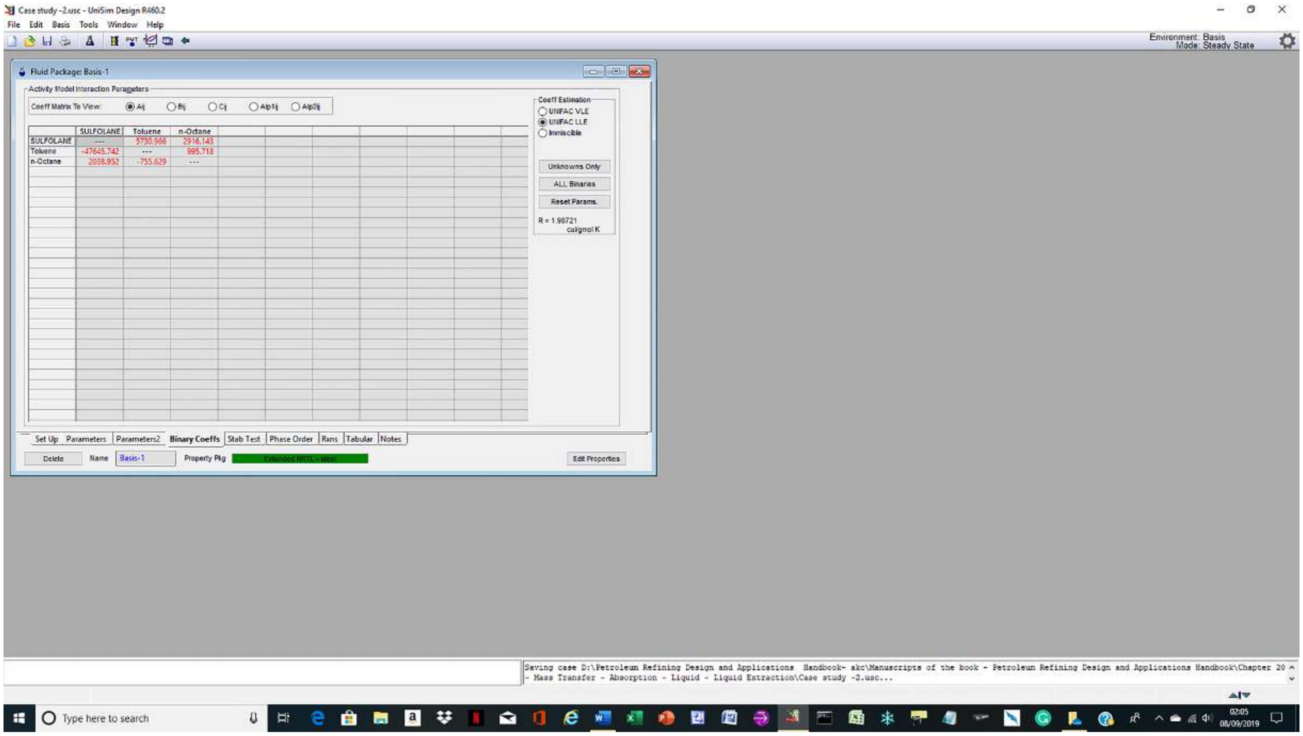
Fluid Pkg for New Sub-FlowSheets

● Use Default Fluid Pkg ☐ Include Column

○ Use Parent's Fluid Pkg

```
Saving case D:\Petroleum Refining Design and Applications  Handbook- akc\Manuscripts of the book - Petroleum Refining Design and Applications Handbook\Chapter 20
- Mass Transfer - Absorption - Liquid - Liquid Extraction\Case study -2.usc...
```

Figure 20.131 A snap-shot of the Fluid Pkgs page on the Simulation Basis Manager window (courtesy of Honeywell Process Solution, UniSim Design R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



Activity Model Interaction Parameters

Coeff Matrix To View: ● Aij ○ Bij ○ Cij ○ Alp1ij ○ Alp2ij

	SULFOLANE	Toluene	n-Octane
SULFOLANE	---	5730.966	2916.143
Toluene	-47645.742	---	995.718
n-Octane	2038.952	-755.629	---

Coeff Estimation

○ UNIFAC VLE

● UNIFAC LLE

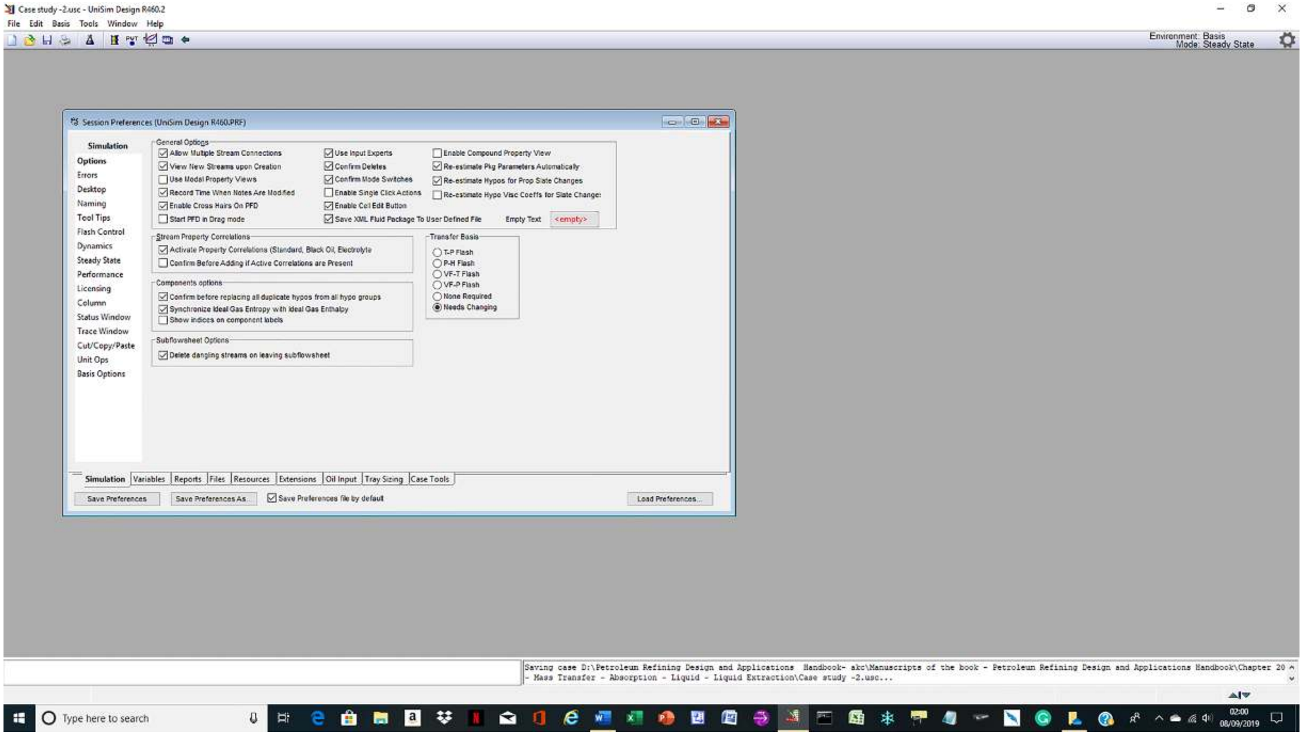
○ Immiscible

R = 1.98721 cal/gmol K

Name: Basis-1 Property Pkg: Extended NRTL - Ideal

```
Saving case D:\Petroleum Refining Design and Applications  Handbook- akc\Manuscripts of the book - Petroleum Refining Design and Applications Handbook\Chapter 20
- Mass Transfer - Absorption - Liquid - Liquid Extraction\Case study -2.usc...
```

Figure 20.132 A snap-shot of the Activity Model Interaction Parameters Window showing the Binary Coefficients of the components (courtesy of Honeywell Process Solution, UniSim Design R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



General Options

- ☑ Allow Multiple Stream Connections
- ☑ View New Streams upon Creation
- ☐ Use Modal Property Views
- ☑ Record Time When Notes Are Modified
- ☑ Enable Cross Hairs On PFD
- ☐ Start PFD in Drag mode
- ☑ Use Input Experts
- ☑ Confirm Deletes
- ☑ Confirm Mode Switches
- ☐ Enable Single Click Actions
- ☑ Enable Cell Edit Button
- ☑ Save XML Fluid Package To User Defined File
- ☐ Enable Compound Property View
- ☑ Re-estimate Pkg Parameters Automatically
- ☑ Re-estimate Hypos for Prop State Changes
- ☐ Re-estimate Hypo Visc Coeffs for State Change:

Empty Text <empty>

Stream Property Correlations

- ☑ Activate Property Correlations (Standard, Black Oil, Electrolyte
- ☐ Confirm Before Adding if Active Correlations are Present

Components options

- ☑ Confirm before replacing all duplicate hypos from all hypo groups
- ☑ Synchronize Ideal Gas Entropy with Ideal Gas Enthalpy
- ☐ Show indices on component labels

Subflowsheet Options

- ☑ Delete dangling streams on leaving subflowsheet

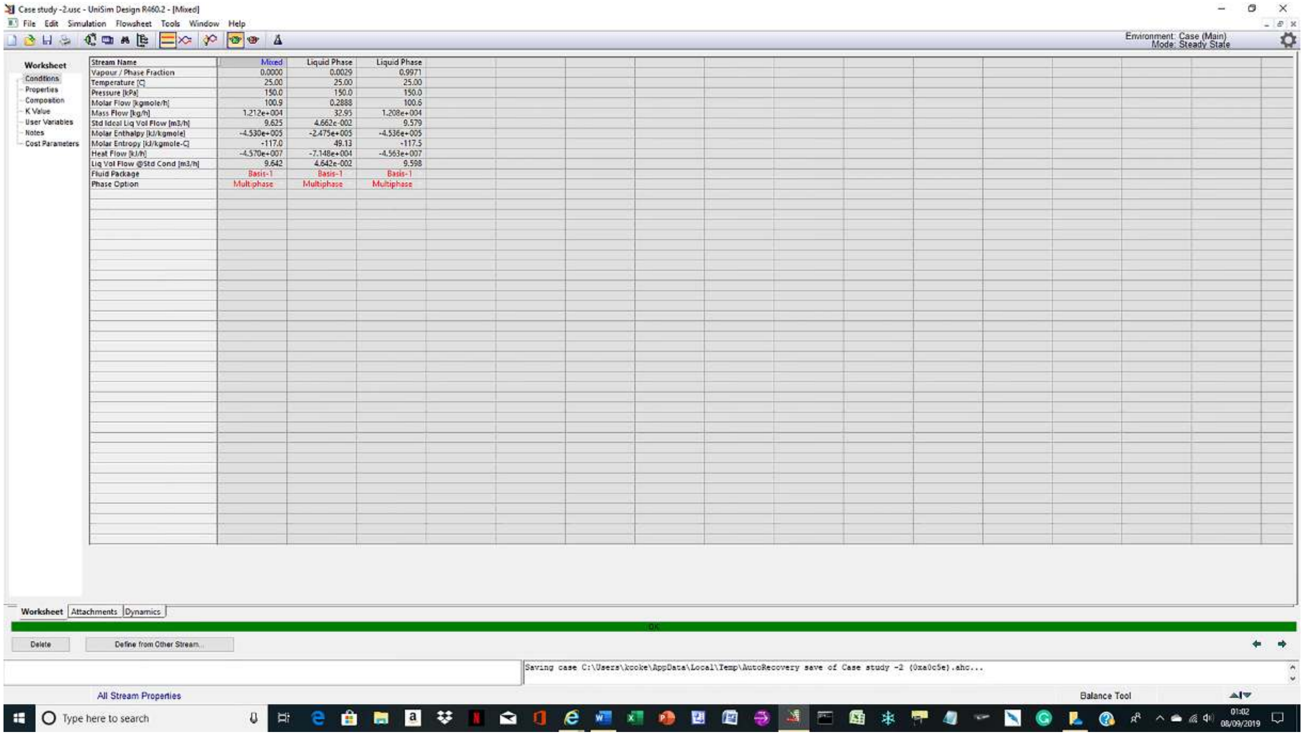
Transfer Basis

- ○ T-P Flash
- ○ P-H Flash
- ○ VF-T Flash
- ○ VF-P Flash
- ○ None Required
- ● Needs Changing

☑ Save Preferences file by default

```
Saving case D:\Petroleum Refining Design and Applications  Handbook- akc\Manuscripts of the book - Petroleum Refining Design and Applications Handbook\Chapter 20
- Mass Transfer - Absorption - Liquid - Liquid Extraction\Case study -2.usc...
```

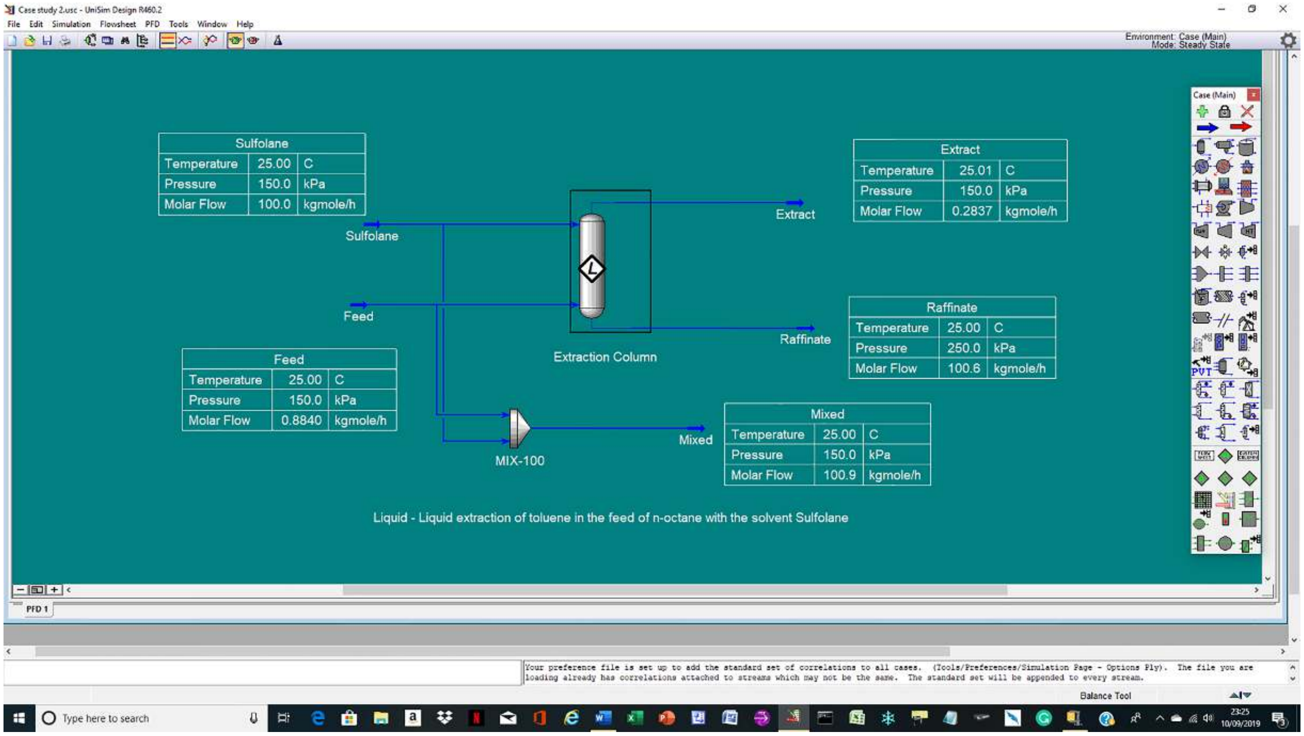
Figure 20.133 A snap – shot of the Session Preferences Window showing General Options with ticked Allow Multiple Stream Connections (courtesy of Honeywell Process Solution, UniSim Design R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



Stream Name	Mixed	Liquid Phase	Liquid Phase
Vapour / Phase Fraction	0.0000	0.0029	0.9971
Temperature [C]	25.00	25.00	25.00
Pressure [kPa]	150.0	150.0	150.0
Molar Flow [kgmole/h]	100.9	0.2888	100.6
Mass Flow [kg/h]	1.212e+004	32.95	1.208e+004
Std Ideal Liq Vol Flow [m3/h]	9.625	4.662e-002	9.579
Molar Enthalpy [kJ/kgmole]	-4.530e+005	-2.475e+005	-4.536e+005
Molar Entropy [kJ/kgmole-C]	-117.0	49.13	-117.5
Heat Flow [kJ/h]	-4.570e+007	-7.148e+004	-4.563e+007
Liq Vol Flow @Std Cond [m3/h]	9.642	4.642e-002	9.598
Fluid Package	Basis-1	Basis-1	Basis-1
Phase Option	Multiphase	Multiphase	Multiphase

```
Saving case C:\Users\kcoke\AppData\Local\Temp\AutoRecovery save of Case study -2 (0xa0c5e).ahc...
```

Figure 20.134 A snap-shot showing two liquid phases from the BIPs parameters generated in Figure 20.132 (courtesy of Honeywell Process Solution, UniSim Design R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



Sulfolane		
Temperature	25.00	C
Pressure	150.0	kPa
Molar Flow	100.0	kgmole/h

Feed		
Temperature	25.00	C
Pressure	150.0	kPa
Molar Flow	0.8840	kgmole/h

Extract		
Temperature	25.01	C
Pressure	150.0	kPa
Molar Flow	0.2837	kgmole/h

Raffinate		
Temperature	25.00	C
Pressure	250.0	kPa
Molar Flow	100.6	kgmole/h

Mixed		
Temperature	25.00	C
Pressure	150.0	kPa
Molar Flow	100.9	kgmole/h

Liquid - Liquid extraction of toluene in the feed of n-octane with the solvent Sulfolane

```
Your preference file is set up to add the standard set of correlations to all cases.  (Tools/Preferences/Simulation Page - Options Ply).  The file you are
loading already has correlations attached to streams which may not be the same.  The standard set will be appended to every stream.
```

Figure 20.135 A snap-shot of the Process flow diagram of liquid – liquid extraction of the feed (Toluene and n-Octane) and solvent (Sulfolane) with the extract and raffinate phases (courtesy of Honeywell Process Solution, UniSim Design R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).

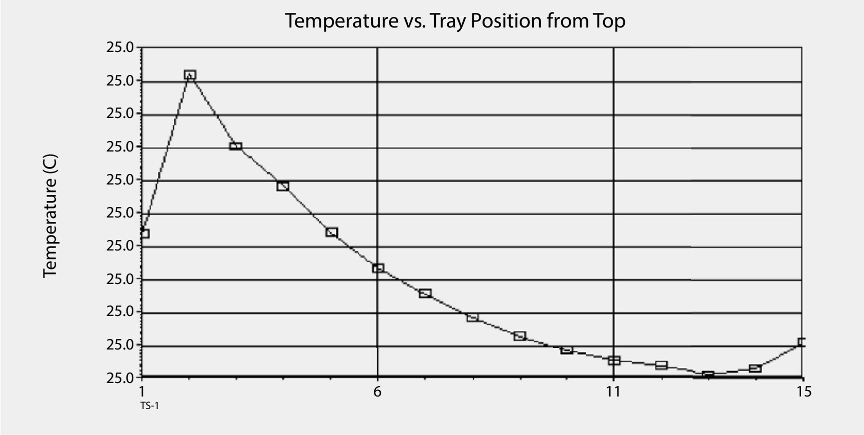
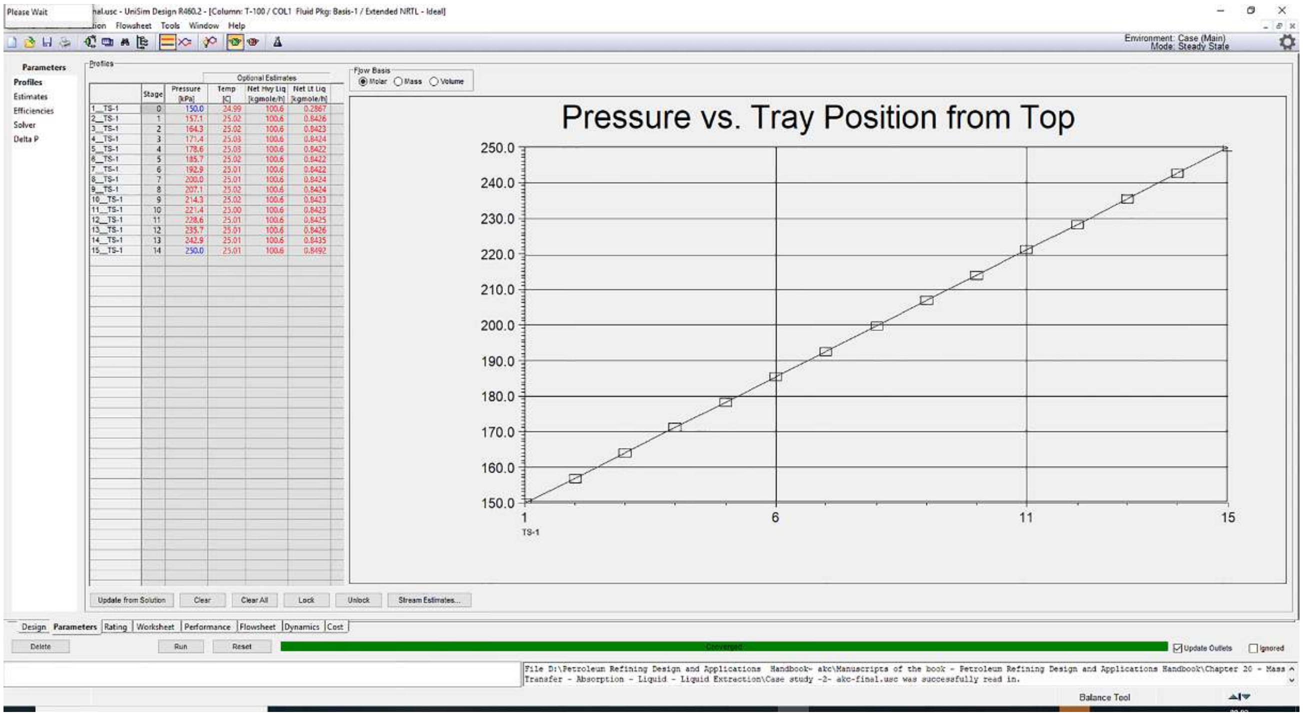


Figure 20.136 Temperature profile vs. tray position from the top of the column (courtesy of Honeywell Process Solution, UniSim Design R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).



	Stage	Pressure [kPa]	Optional Estimates Temp [C]	Net Hvy Liq [kgmole/h]	Net Lt Liq [kgmole/h]
1__TS-1	0	150.0	24.99	100.6	0.2867
2__TS-1	1	157.1	25.02	100.6	0.8426
3__TS-1	2	164.3	25.02	100.6	0.8423
4__TS-1	3	171.4	25.03	100.6	0.8424
5__TS-1	4	178.6	25.03	100.6	0.8422
6__TS-1	5	185.7	25.02	100.6	0.8422
7__TS-1	6	192.9	25.01	100.6	0.8422
8__TS-1	7	200.0	25.01	100.6	0.8424
9__TS-1	8	207.1	25.02	100.6	0.8424
10__TS-1	9	214.3	25.02	100.6	0.8423
11__TS-1	10	221.4	25.00	100.6	0.8423
12__TS-1	11	228.6	25.01	100.6	0.8425
13__TS-1	12	235.7	25.01	100.6	0.8426
14__TS-1	13	242.9	25.01	100.6	0.8435
15__TS-1	14	250.0	25.01	100.6	0.8492

```
File D:\Petroleum Refining Design and Applications  Handbook- akc\Manuscripts of the book - Petroleum Refining Design and Applications Handbook\Chapter 20 - Mass
Transfer - Absorption - Liquid - Liquid Extraction\Case study -2- akc-final.usc was successfully read in.
```

Figure 20.137 A snap-shot of the pressure profile vs. tray position from the top of the column (courtesy of Honeywell Process Solution, UniSim Design R460.2, Honeywell®, and UniSim® are registered trademarks of Honeywell International Inc.).

Feeds:

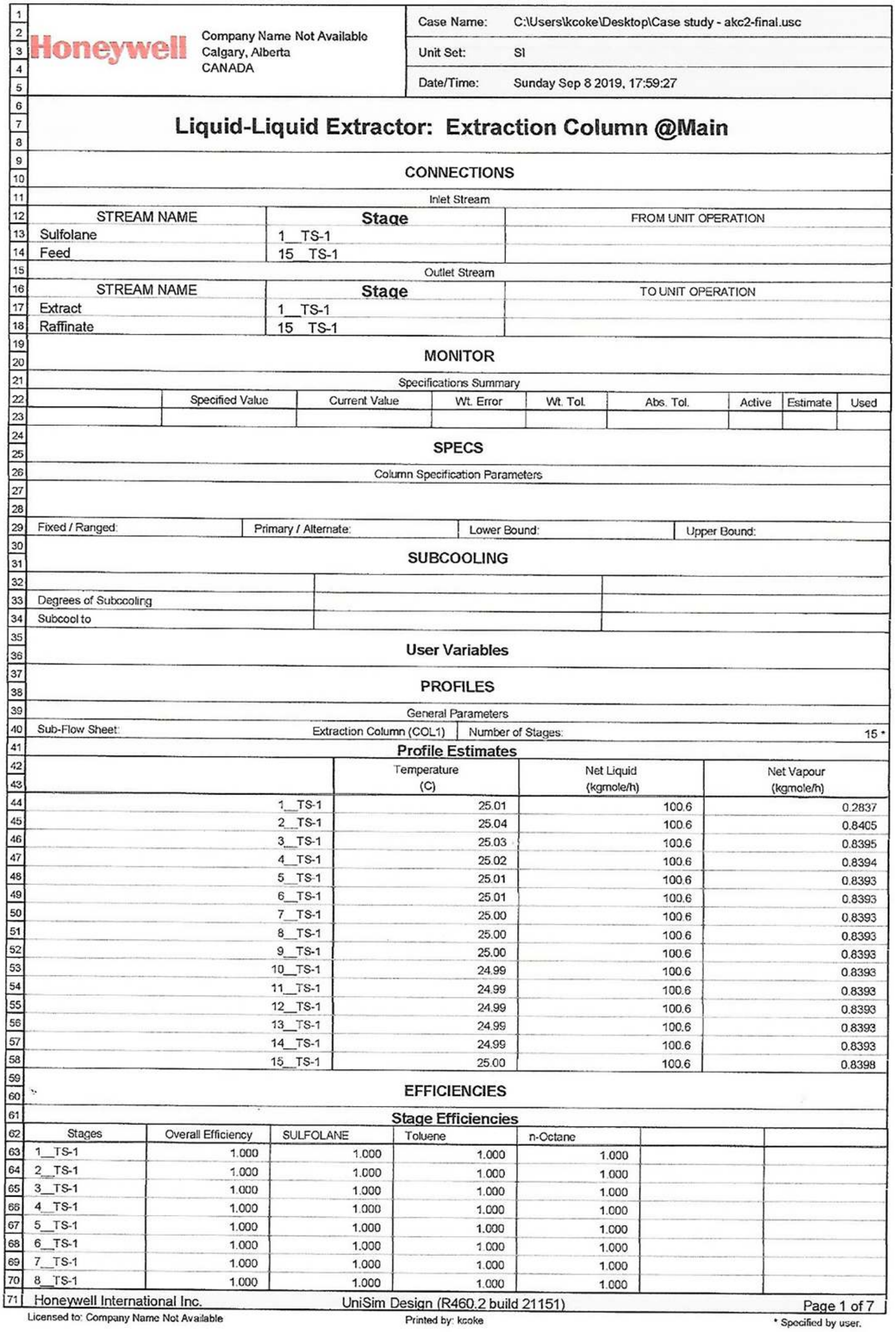
	Sulfolane	**Feed**
Flow Rate (kg mole/h)	100.0	0.8840
Sulfolane	1.00	0.000
Toluene	0.00	0.050
n-Octane	0.00	0.950

Products

	Extract	**Raffinate**
Flow Rate (kg mole/h)	0.2837	100.6003
Sulfolane	0.0014	0.9940
Toluene	0.0000	0.0004
n-Octane	0.9986	0.0055

Tables 20.65 to 20.71 show the results of the simulation using UniSim Design® 460.2.

Table 20.65 Results of liquid–liquid extraction simulation of Case Study-2.



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name:	C:\Users\kcoke\Desktop\Case study - akc2-final.usc
Unit Set:	SI
Date/Time:	Sunday Sep 8 2019, 17:59:27

Liquid-Liquid Extractor: Extraction Column @Main

CONNECTIONS

Inlet Stream

STREAM NAME	Stage	FROM UNIT OPERATION
Sulfolane	1__TS-1	
Feed	15__TS-1	

Outlet Stream

STREAM NAME	Stage	TO UNIT OPERATION
Extract	1__TS-1	
Raffinate	15__TS-1	

MONITOR

Specifications Summary

	Specified Value	Current Value	Wt. Error	Wt. Tol.	Abs. Tol.	Active	Estimate	Used

SPECS

Column Specification Parameters

Fixed / Ranged:	Primary / Alternate:	Lower Bound:	Upper Bound:

SUBCOOLING

Degrees of Subcooling		
Subcool to		

User Variables

PROFILES

General Parameters

Sub-Flow Sheet:	Extraction Column (COL1)	Number of Stages:	15 *

Profile Estimates

	Temperature (C)	Net Liquid (kgmole/h)	Net Vapour (kgmole/h)
1__TS-1	25.01	100.6	0.2837
2__TS-1	25.04	100.6	0.8405
3__TS-1	25.03	100.6	0.8395
4__TS-1	25.02	100.6	0.8394
5__TS-1	25.01	100.6	0.8393
6__TS-1	25.01	100.6	0.8393
7__TS-1	25.00	100.6	0.8393
8__TS-1	25.00	100.6	0.8393
9__TS-1	25.00	100.6	0.8393
10__TS-1	24.99	100.6	0.8393
11__TS-1	24.99	100.6	0.8393
12__TS-1	24.99	100.6	0.8393
13__TS-1	24.99	100.6	0.8393
14__TS-1	24.99	100.6	0.8393
15__TS-1	25.00	100.6	0.8398

EFFICIENCIES

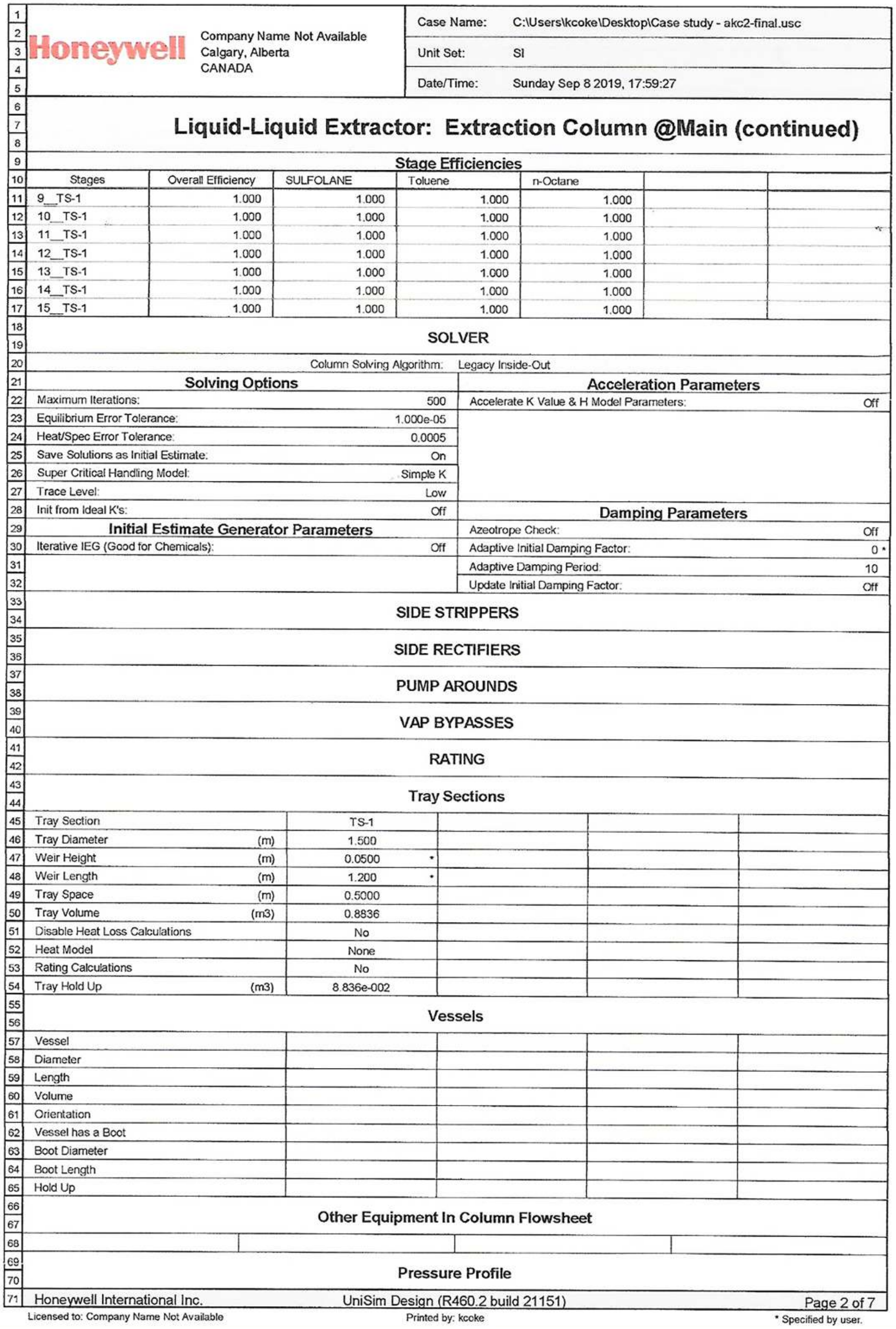
Stage Efficiencies

Stages	Overall Efficiency	SULFOLANE	Toluene	n-Octane		
1__TS-1	1.000	1.000	1.000	1.000		
2__TS-1	1.000	1.000	1.000	1.000		
3__TS-1	1.000	1.000	1.000	1.000		
4__TS-1	1.000	1.000	1.000	1.000		
5__TS-1	1.000	1.000	1.000	1.000		
6__TS-1	1.000	1.000	1.000	1.000		
7__TS-1	1.000	1.000	1.000	1.000		
8__TS-1	1.000	1.000	1.000	1.000		

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 1 of 7

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

Table 20.66 Results of liquid–liquid extraction simulation of Case Study-2. (*Continued*)



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name: C:\Users\kcoke\Desktop\Case study - akc2-final.usc
Unit Set: SI
Date/Time: Sunday Sep 8 2019, 17:59:27

Liquid-Liquid Extractor: Extraction Column @Main (continued)

Stage Efficiencies

Stages	Overall Efficiency	SULFOLANE	Toluene	n-Octane		
9__TS-1	1.000	1.000	1.000	1.000		
10__TS-1	1.000	1.000	1.000	1.000		
11__TS-1	1.000	1.000	1.000	1.000		
12__TS-1	1.000	1.000	1.000	1.000		
13__TS-1	1.000	1.000	1.000	1.000		
14__TS-1	1.000	1.000	1.000	1.000		
15__TS-1	1.000	1.000	1.000	1.000		

SOLVER

Column Solving Algorithm: Legacy Inside-Out

Solving Options

Maximum Iterations:	500
Equilibrium Error Tolerance:	1.000e-05
Heat/Spec Error Tolerance:	0.0005
Save Solutions as Initial Estimate:	On
Super Critical Handling Model:	Simple K
Trace Level:	Low
Init from Ideal K's:	Off

Initial Estimate Generator Parameters

Iterative IEG (Good for Chemicals):	Off

Acceleration Parameters

Accelerate K Value & H Model Parameters:	Off

Damping Parameters

Azeotrope Check:	Off
Adaptive Initial Damping Factor:	0 *
Adaptive Damping Period:	10
Update Initial Damping Factor:	Off

SIDE STRIPPERS

SIDE RECTIFIERS

PUMP AROUNDS

VAP BYPASSES

RATING

Tray Sections

Tray Section		TS-1			
Tray Diameter	(m)	1.500			
Weir Height	(m)	0.0500 *			
Weir Length	(m)	1.200 *			
Tray Space	(m)	0.5000			
Tray Volume	(m3)	0.8836			
Disable Heat Loss Calculations		No			
Heat Model		None			
Rating Calculations		No			
Tray Hold Up	(m3)	8.836e-002			

Vessels

Vessel				
Diameter				
Length				
Volume				
Orientation				
Vessel has a Boot				
Boot Diameter				
Boot Length				
Hold Up				

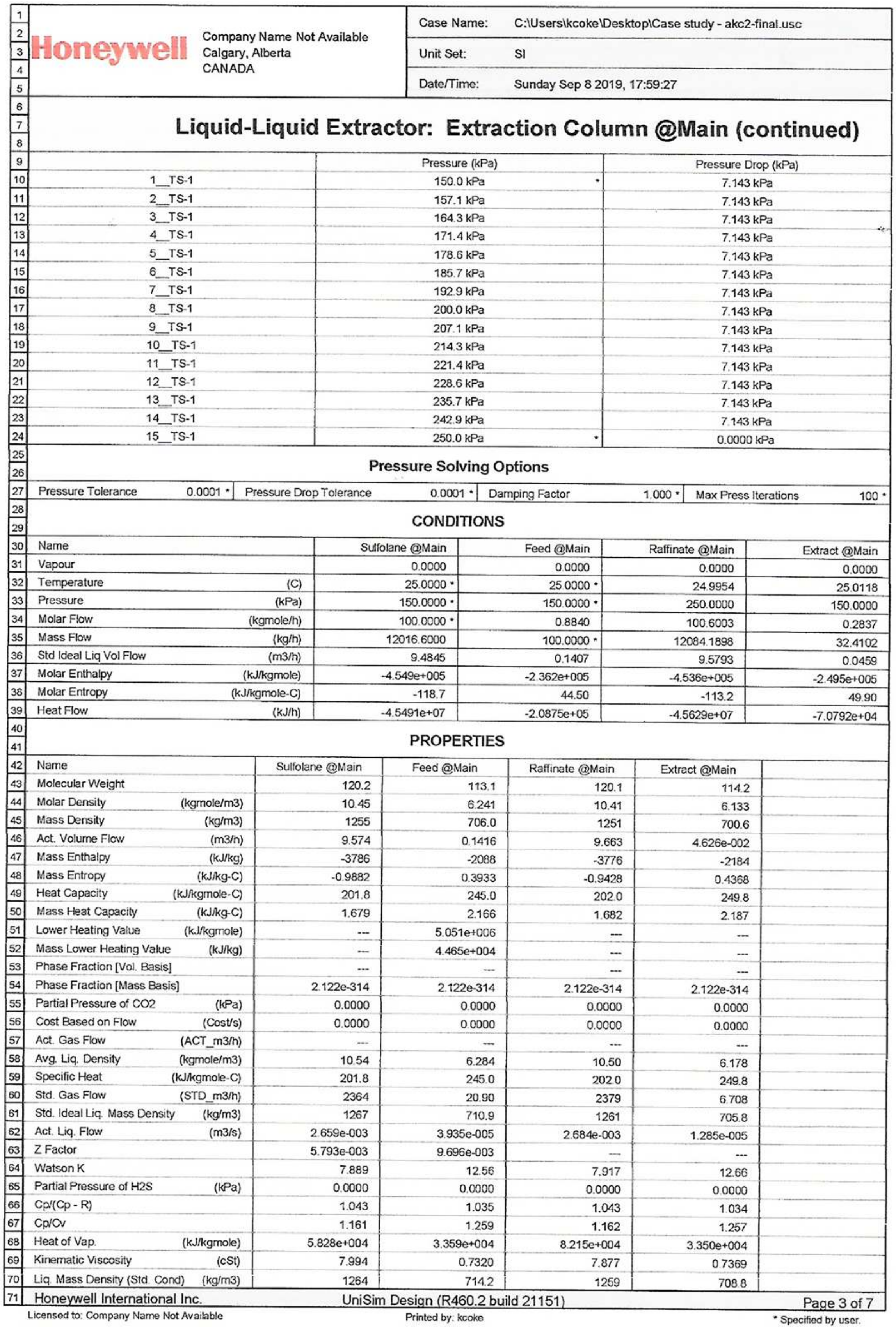
Other Equipment In Column Flowsheet

Pressure Profile

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 2 of 7

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

Table 20.67 Results of liquid–liquid extraction simulation of Case Study-2. (*Continued*)



Honeywell	Company Name Not Available Calgary, Alberta CANADA	Case Name: C:\Users\kcoke\Desktop\Case study - akc2-final.usc Unit Set: SI Date/Time: Sunday Sep 8 2019, 17:59:27

Liquid-Liquid Extractor: Extraction Column @Main (continued)

	Pressure (kPa)	Pressure Drop (kPa)
1__TS-1	150.0 kPa *	7.143 kPa
2__TS-1	157.1 kPa	7.143 kPa
3__TS-1	164.3 kPa	7.143 kPa
4__TS-1	171.4 kPa	7.143 kPa
5__TS-1	178.6 kPa	7.143 kPa
6__TS-1	185.7 kPa	7.143 kPa
7__TS-1	192.9 kPa	7.143 kPa
8__TS-1	200.0 kPa	7.143 kPa
9__TS-1	207.1 kPa	7.143 kPa
10__TS-1	214.3 kPa	7.143 kPa
11__TS-1	221.4 kPa	7.143 kPa
12__TS-1	228.6 kPa	7.143 kPa
13__TS-1	235.7 kPa	7.143 kPa
14__TS-1	242.9 kPa	7.143 kPa
15__TS-1	250.0 kPa *	0.0000 kPa

Pressure Solving Options

Pressure Tolerance	0.0001 *	Pressure Drop Tolerance	0.0001 *	Damping Factor	1.000 *	Max Press Iterations	100 *

CONDITIONS

Name		Sulfolane @Main	Feed @Main	Raffinate @Main	Extract @Main
Vapour		0.0000	0.0000	0.0000	0.0000
Temperature	(C)	25.0000 *	25.0000 *	24.9954	25.0118
Pressure	(kPa)	150.0000 *	150.0000 *	250.0000	150.0000
Molar Flow	(kgmole/h)	100.0000 *	0.8840	100.6003	0.2837
Mass Flow	(kg/h)	12016.6000	100.0000 *	12084.1898	32.4102
Std Ideal Liq Vol Flow	(m3/h)	9.4845	0.1407	9.5793	0.0459
Molar Enthalpy	(kJ/kgmole)	-4.549e+005	-2.362e+005	-4.536e+005	-2.495e+005
Molar Entropy	(kJ/kgmole-C)	-118.7	44.50	-113.2	49.90
Heat Flow	(kJ/h)	-4.5491e+07	-2.0875e+05	-4.5629e+07	-7.0792e+04

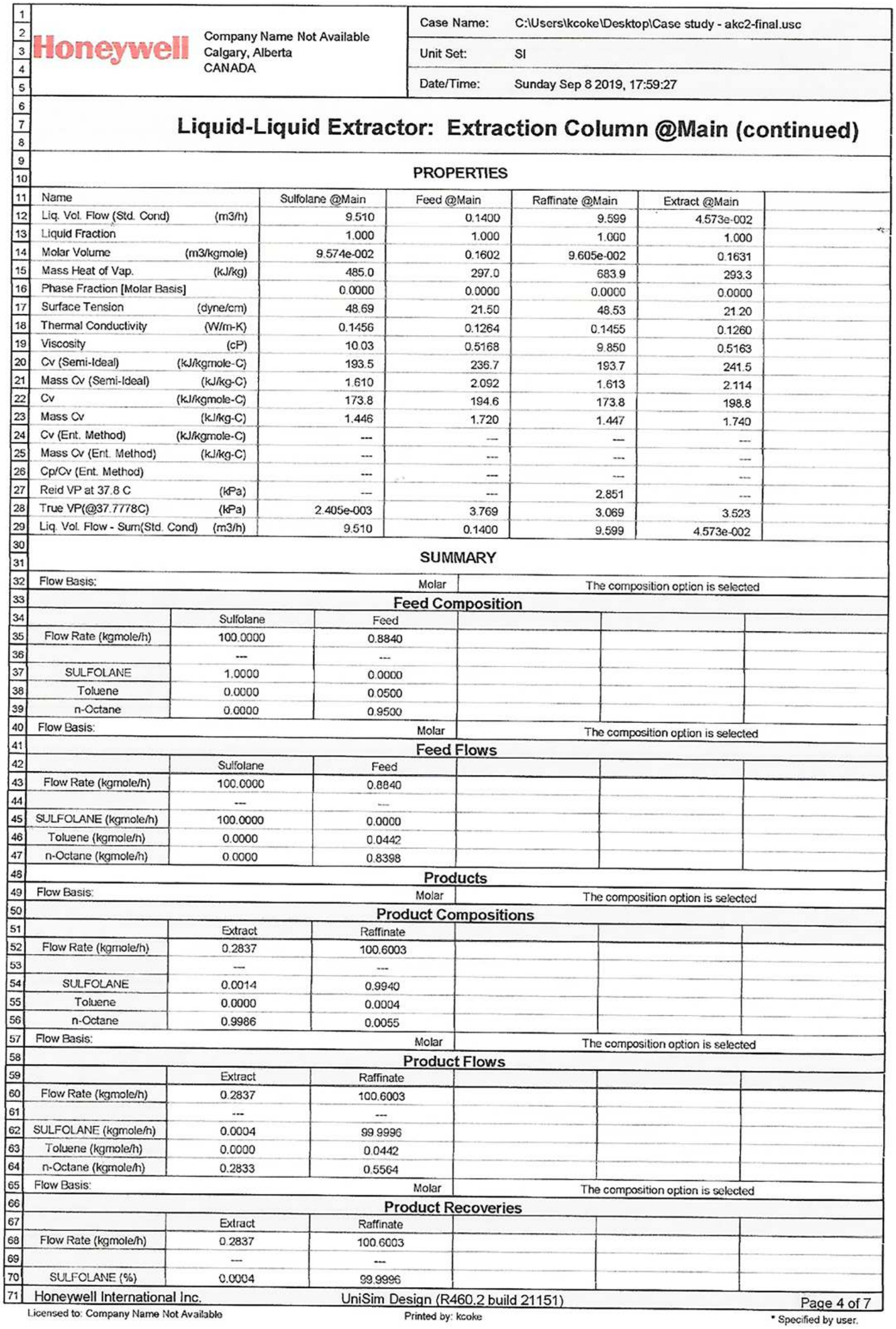
PROPERTIES

Name		Sulfolane @Main	Feed @Main	Raffinate @Main	Extract @Main
Molecular Weight		120.2	113.1	120.1	114.2
Molar Density	(kgmole/m3)	10.45	6.241	10.41	6.133
Mass Density	(kg/m3)	1255	706.0	1251	700.6
Act. Volume Flow	(m3/h)	9.574	0.1416	9.663	4.626e-002
Mass Enthalpy	(kJ/kg)	-3786	-2088	-3776	-2184
Mass Entropy	(kJ/kg-C)	-0.9882	0.3933	-0.9428	0.4368
Heat Capacity	(kJ/kgmole-C)	201.8	245.0	202.0	249.8
Mass Heat Capacity	(kJ/kg-C)	1.679	2.166	1.682	2.187
Lower Heating Value	(kJ/kgmole)	---	5.051e+006	---	---
Mass Lower Heating Value	(kJ/kg)	---	4.465e+004	---	---
Phase Fraction [Vol. Basis]		---	---	---	---
Phase Fraction [Mass Basis]		2.122e-314	2.122e-314	2.122e-314	2.122e-314
Partial Pressure of CO2	(kPa)	0.0000	0.0000	0.0000	0.0000
Cost Based on Flow	(Cost/s)	0.0000	0.0000	0.0000	0.0000
Act. Gas Flow	(ACT_m3/h)	---	---	---	---
Avg. Liq. Density	(kgmole/m3)	10.54	6.284	10.50	6.178
Specific Heat	(kJ/kgmole-C)	201.8	245.0	202.0	249.8
Std. Gas Flow	(STD_m3/h)	2364	20.90	2379	6.708
Std. Ideal Liq. Mass Density	(kg/m3)	1267	710.9	1261	705.8
Act. Liq. Flow	(m3/s)	2.659e-003	3.935e-005	2.684e-003	1.285e-005
Z Factor		5.793e-003	9.696e-003	---	---
Watson K		7.889	12.56	7.917	12.66
Partial Pressure of H2S	(kPa)	0.0000	0.0000	0.0000	0.0000
Cp/(Cp - R)		1.043	1.035	1.043	1.034
Cp/Cv		1.161	1.259	1.162	1.257
Heat of Vap.	(kJ/kgmole)	5.828e+004	3.359e+004	8.215e+004	3.350e+004
Kinematic Viscosity	(cSt)	7.994	0.7320	7.877	0.7369
Liq. Mass Density (Std. Cond)	(kg/m3)	1264	714.2	1259	708.8

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 3 of 7

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

Table 20.68 Results of liquid–liquid extraction simulation of Case Study-2. (*Continued*)



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name: C:\Users\kcoke\Desktop\Case study - akc2-final.usc
Unit Set: SI
Date/Time: Sunday Sep 8 2019, 17:59:27

Liquid-Liquid Extractor: Extraction Column @Main (continued)

PROPERTIES

Name	Sulfolane @Main	Feed @Main	Raffinate @Main	Extract @Main
Liq. Vol. Flow (Std. Cond) (m3/h)	9.510	0.1400	9.599	4.573e-002
Liquid Fraction	1.000	1.000	1.000	1.000
Molar Volume (m3/kgmole)	9.574e-002	0.1602	9.605e-002	0.1631
Mass Heat of Vap. (kJ/kg)	485.0	297.0	683.9	293.3
Phase Fraction [Molar Basis]	0.0000	0.0000	0.0000	0.0000
Surface Tension (dyne/cm)	48.69	21.50	48.53	21.20
Thermal Conductivity (W/m-K)	0.1456	0.1264	0.1455	0.1260
Viscosity (cP)	10.03	0.5168	9.850	0.5163
Cv (Semi-Ideal) (kJ/kgmole-C)	193.5	236.7	193.7	241.5
Mass Cv (Semi-Ideal) (kJ/kg-C)	1.610	2.092	1.613	2.114
Cv (kJ/kgmole-C)	173.8	194.6	173.8	198.8
Mass Cv (kJ/kg-C)	1.446	1.720	1.447	1.740
Cv (Ent. Method) (kJ/kgmole-C)	---	---	---	---
Mass Cv (Ent. Method) (kJ/kg-C)	---	---	---	---
Cp/Cv (Ent. Method)	---	---	---	---
Reid VP at 37.8 C (kPa)	---	---	2.851	---
True VP(@37.7778C) (kPa)	2.405e-003	3.769	3.069	3.523
Liq. Vol. Flow - Sum(Std. Cond) (m3/h)	9.510	0.1400	9.599	4.573e-002

SUMMARY

Flow Basis: Molar The composition option is selected

Feed Composition

	Sulfolane	Feed
Flow Rate (kgmole/h)	100.0000	0.8840
	---	---
SULFOLANE	1.0000	0.0000
Toluene	0.0000	0.0500
n-Octane	0.0000	0.9500

Flow Basis: Molar The composition option is selected

Feed Flows

	Sulfolane	Feed
Flow Rate (kgmole/h)	100.0000	0.8840
	---	---
SULFOLANE (kgmole/h)	100.0000	0.0000
Toluene (kgmole/h)	0.0000	0.0442
n-Octane (kgmole/h)	0.0000	0.8398

Products

Flow Basis: Molar The composition option is selected

Product Compositions

	Extract	Raffinate
Flow Rate (kgmole/h)	0.2837	100.6003
	---	---
SULFOLANE	0.0014	0.9940
Toluene	0.0000	0.0004
n-Octane	0.9986	0.0055

Flow Basis: Molar The composition option is selected

Product Flows

	Extract	Raffinate
Flow Rate (kgmole/h)	0.2837	100.6003
	---	---
SULFOLANE (kgmole/h)	0.0004	99.9996
Toluene (kgmole/h)	0.0000	0.0442
n-Octane (kgmole/h)	0.2833	0.5564

Flow Basis: Molar The composition option is selected

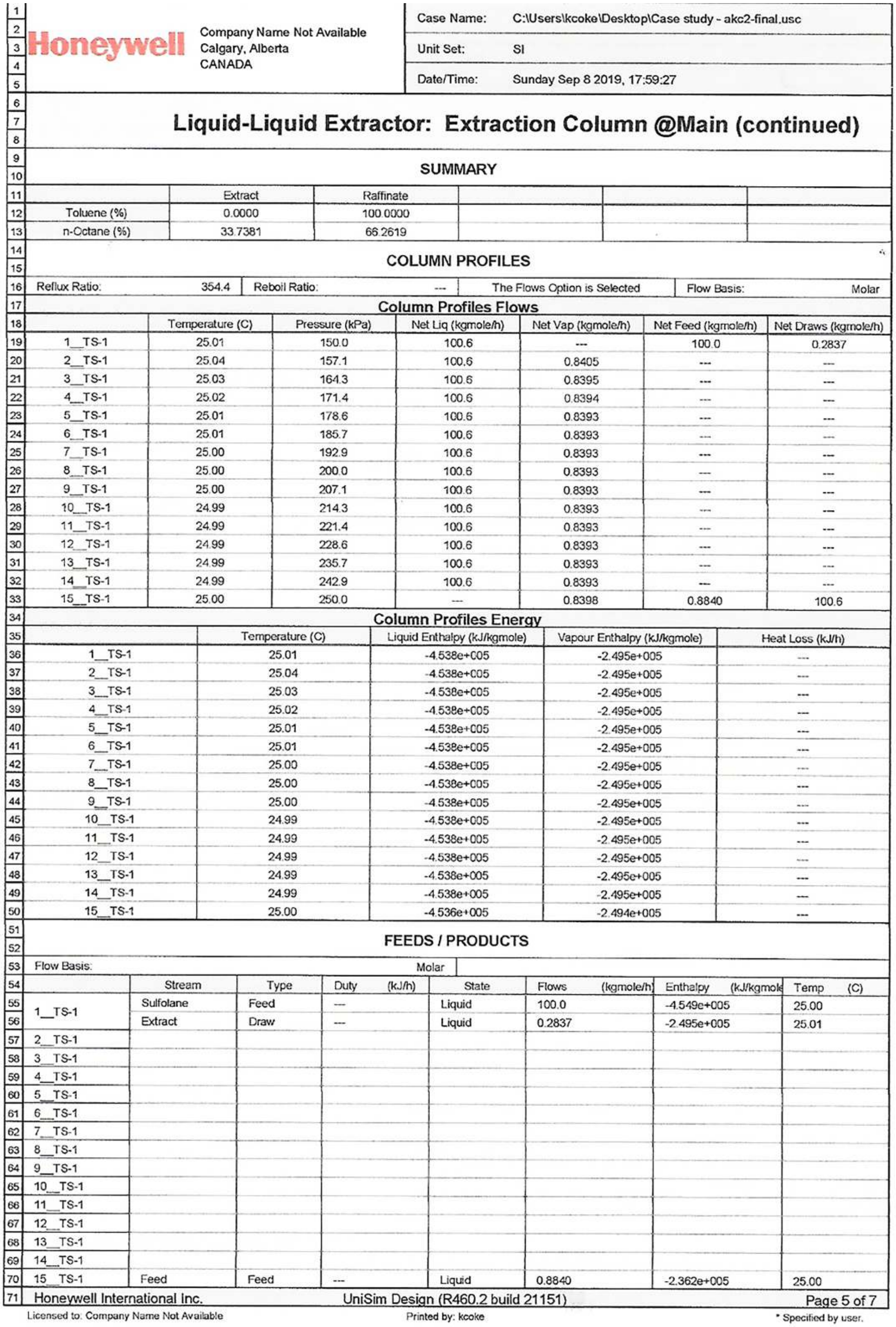
Product Recoveries

	Extract	Raffinate
Flow Rate (kgmole/h)	0.2837	100.6003
	---	---
SULFOLANE (%)	0.0004	99.9996

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 4 of 7

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

Table 20.69 Results of liquid–liquid extraction simulation of Case Study-2. (*Continued*)



Honeywell

Company Name Not Available
Calgary, Alberta
CANADA

Case Name: C:\Users\kcoke\Desktop\Case study - akc2-final.usc
Unit Set: SI
Date/Time: Sunday Sep 8 2019, 17:59:27

Liquid-Liquid Extractor: Extraction Column @Main (continued)

SUMMARY

	Extract	Raffinate			
Toluene (%)	0.0000	100.0000			
n-Octane (%)	33.7381	66.2619			

COLUMN PROFILES

Reflux Ratio:	354.4	Reboil Ratio:	---	The Flows Option is Selected	Flow Basis: Molar

Column Profiles Flows

	Temperature (C)	Pressure (kPa)	Net Liq (kgmole/h)	Net Vap (kgmole/h)	Net Feed (kgmole/h)	Net Draws (kgmole/h)
1__TS-1	25.01	150.0	100.6	---	100.0	0.2837
2__TS-1	25.04	157.1	100.6	0.8405	---	---
3__TS-1	25.03	164.3	100.6	0.8395	---	---
4__TS-1	25.02	171.4	100.6	0.8394	---	---
5__TS-1	25.01	178.6	100.6	0.8393	---	---
6__TS-1	25.01	185.7	100.6	0.8393	---	---
7__TS-1	25.00	192.9	100.6	0.8393	---	---
8__TS-1	25.00	200.0	100.6	0.8393	---	---
9__TS-1	25.00	207.1	100.6	0.8393	---	---
10__TS-1	24.99	214.3	100.6	0.8393	---	---
11__TS-1	24.99	221.4	100.6	0.8393	---	---
12__TS-1	24.99	228.6	100.6	0.8393	---	---
13__TS-1	24.99	235.7	100.6	0.8393	---	---
14__TS-1	24.99	242.9	100.6	0.8393	---	---
15__TS-1	25.00	250.0	---	0.8398	0.8840	100.6

Column Profiles Energy

	Temperature (C)	Liquid Enthalpy (kJ/kgmole)	Vapour Enthalpy (kJ/kgmole)	Heat Loss (kJ/h)
1__TS-1	25.01	-4.538e+005	-2.495e+005	---
2__TS-1	25.04	-4.538e+005	-2.495e+005	---
3__TS-1	25.03	-4.538e+005	-2.495e+005	---
4__TS-1	25.02	-4.538e+005	-2.495e+005	---
5__TS-1	25.01	-4.538e+005	-2.495e+005	---
6__TS-1	25.01	-4.538e+005	-2.495e+005	---
7__TS-1	25.00	-4.538e+005	-2.495e+005	---
8__TS-1	25.00	-4.538e+005	-2.495e+005	---
9__TS-1	25.00	-4.538e+005	-2.495e+005	---
10__TS-1	24.99	-4.538e+005	-2.495e+005	---
11__TS-1	24.99	-4.538e+005	-2.495e+005	---
12__TS-1	24.99	-4.538e+005	-2.495e+005	---
13__TS-1	24.99	-4.538e+005	-2.495e+005	---
14__TS-1	24.99	-4.538e+005	-2.495e+005	---
15__TS-1	25.00	-4.536e+005	-2.494e+005	---

FEEDS / PRODUCTS

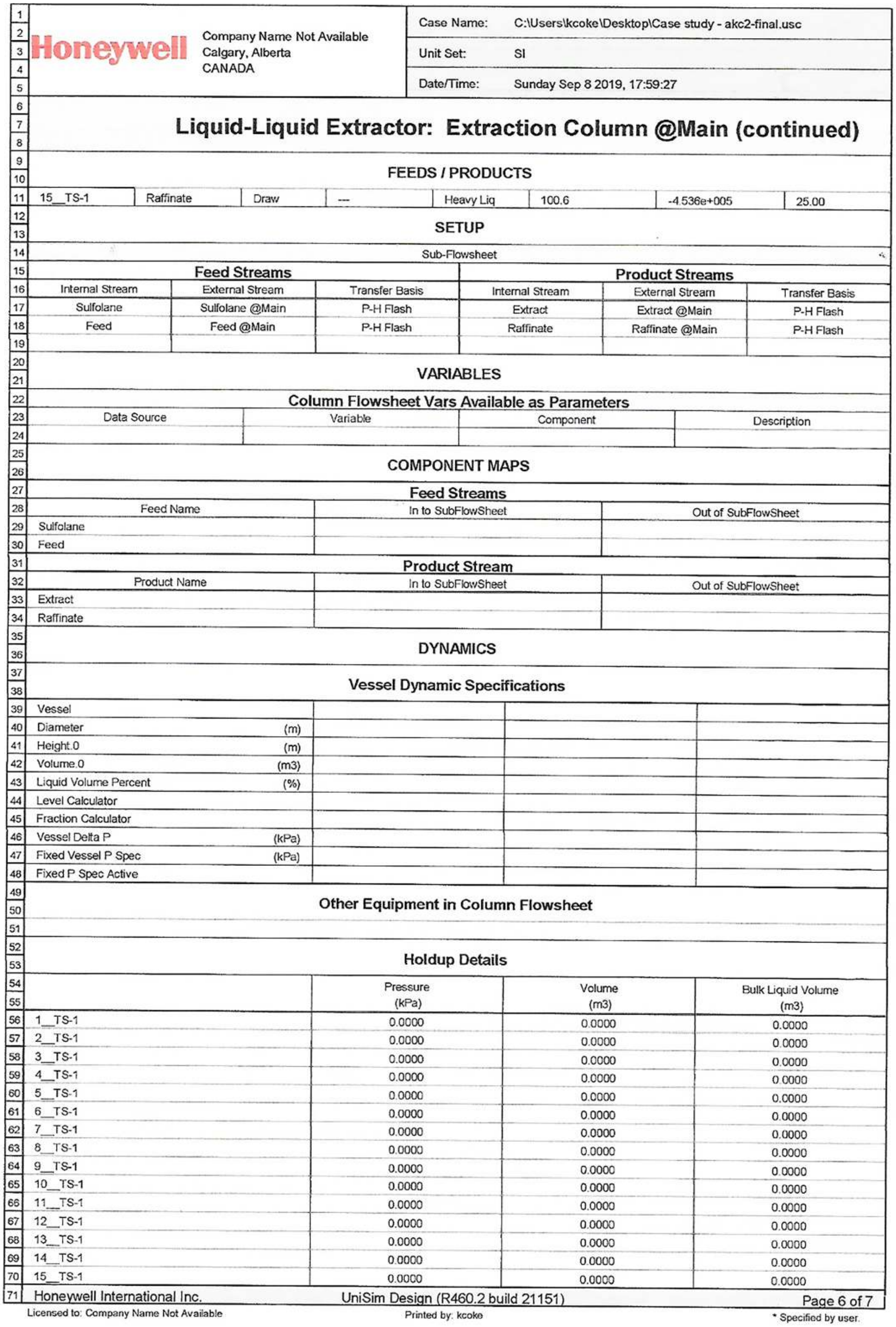
Flow Basis: Molar

	Stream	Type	Duty (kJ/h)	State	Flows (kgmole/h)	Enthalpy (kJ/kgmole)	Temp (C)
1__TS-1	Sulfolane	Feed	---	Liquid	100.0	-4.549e+005	25.00
	Extract	Draw	---	Liquid	0.2837	-2.495e+005	25.01
2__TS-1							
3__TS-1							
4__TS-1							
5__TS-1							
6__TS-1							
7__TS-1							
8__TS-1							
9__TS-1							
10__TS-1							
11__TS-1							
12__TS-1							
13__TS-1							
14__TS-1							
15__TS-1	Feed	Feed	---	Liquid	0.8840	-2.362e+005	25.00

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 5 of 7

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

Table 20.70 Results of liquid–liquid extraction simulation of Case Study-2. (*Continued*)



Honeywell
Company Name Not Available
Calgary, Alberta
CANADA

Case Name: C:\Users\kcoke\Desktop\Case study - akc2-final.usc
Unit Set: SI
Date/Time: Sunday Sep 8 2019, 17:59:27

Liquid-Liquid Extractor: Extraction Column @Main (continued)

FEEDS / PRODUCTS

15__TS-1	Raffinate	Draw	---	Heavy Liq	100.6	-4.536e+005	25.00

SETUP

Sub-Flowsheet

Feed Streams			Product Streams		
Internal Stream	External Stream	Transfer Basis	Internal Stream	External Stream	Transfer Basis
Sulfolane	Sulfolane @Main	P-H Flash	Extract	Extract @Main	P-H Flash
Feed	Feed @Main	P-H Flash	Raffinate	Raffinate @Main	P-H Flash

VARIABLES

Column Flowsheet Vars Available as Parameters

Data Source	Variable	Component	Description

COMPONENT MAPS

Feed Streams

Feed Name	In to SubFlowSheet	Out of SubFlowSheet
Sulfolane		
Feed		

Product Stream

Product Name	In to SubFlowSheet	Out of SubFlowSheet
Extract		
Raffinate		

DYNAMICS

Vessel Dynamic Specifications

Vessel			
Diameter (m)			
Height.0 (m)			
Volume.0 (m3)			
Liquid Volume Percent (%)			
Level Calculator			
Fraction Calculator			
Vessel Delta P (kPa)			
Fixed Vessel P Spec (kPa)			
Fixed P Spec Active			

Other Equipment in Column Flowsheet

Holdup Details

	Pressure (kPa)	Volume (m3)	Bulk Liquid Volume (m3)
1__TS-1	0.0000	0.0000	0.0000
2__TS-1	0.0000	0.0000	0.0000
3__TS-1	0.0000	0.0000	0.0000
4__TS-1	0.0000	0.0000	0.0000
5__TS-1	0.0000	0.0000	0.0000
6__TS-1	0.0000	0.0000	0.0000
7__TS-1	0.0000	0.0000	0.0000
8__TS-1	0.0000	0.0000	0.0000
9__TS-1	0.0000	0.0000	0.0000
10__TS-1	0.0000	0.0000	0.0000
11__TS-1	0.0000	0.0000	0.0000
12__TS-1	0.0000	0.0000	0.0000
13__TS-1	0.0000	0.0000	0.0000
14__TS-1	0.0000	0.0000	0.0000
15__TS-1	0.0000	0.0000	0.0000

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 6 of 7
Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

Table 20.71 Results of liquid–liquid extraction simulation of Case Study-2. (*Continued*)

Honeywell	Company Name Not Available Calgary, Alberta CANADA	Case Name: C:\Users\kcoke\Desktop\Case study - akc2-final.usc Unit Set: SI Date/Time: Sunday Sep 8 2019, 17:59:27

Liquid-Liquid Extractor: Extraction Column @Main (continued)

STATUS

OK

NOTES

Description

Honeywell International Inc. UniSim Design (R460.2 build 21151) Page 7 of 7

Licensed to: Company Name Not Available Printed by: kcoke * Specified by user.

References

1. Hoek, P. J. “Large- and Small-Scale Liquid Maldistribution in Packed Column,” PhD, Thesis, Delft University, Netherlands (1983).
2. Strigle, R. F., Jr, Random Packings and Packed Towers, Gulf Publishing Co. Houston, Texas (1987).
3. Eckert, J. S., U.S. Stoneware Co., private communication
4. Bonilla, J. A., “Don’t Neglect Liquid Distributors,” Chem. Eng. Prog., V. 89, No. 3, p. 47 (1993).
5. Kister, H. Z., K. F. Larson and T. Yangi, “How Do Trays and Packing Stack Up?” Chem. Eng. Prog., V. 90, No. 2, p. 23 (1994).
6. Kister, H. Z., K. F. Larson and T. Yangi, “Capacity and Efficiency: How Trays and Packings Compare,” AIChE. Spring Meeting, Mar. (1995).
7. Capps, R. W., “Consider the Ultimate Capacity of Fractionation Trays,” Chem. Eng. Prog., V. 89, No. 3, p. 37 (1993).
8. Chen, G. K. “Packed Column Internals,” Chem. Eng., p. 40. Mar. 5, (1984).
9. Eckert, J. S., “Design Techniques for Sizing Packed Towers,” Chem. Eng. Prog. V. 57, No. 9, p. 54 (1961).
10. Kunesh, J. G., “Practical Tips on Tower Packing,” Chem. Eng., V. 94, No. 18, p. 101 (1987).
11. Billet, R., “Gauze-Packed Columns for Vacuum Distillation,” Chem. Eng., V. 79, No. 4, p. 68 (1972).
12. Perry, R. H. and D. Green, Perry’s Chemical Engineering Handbook, 6th Ed. McGraw-Hill, Inc., New York, N.Y. (1984).

13. Eckert, J. S., "Selecting the Proper Distillation Column Packing," Chem. Eng. Prog., Vol. 66, No. 3, p. 39, March (1970).
14. Leva, M. J. "Tower Packing and Packed Tower Design," 2nd. Ed., U.S. Stoneware Co., Akron, OH.
15. Bulletin TP-54, Tower Packings, U.S. Stoneware Co. Akron, Ohio.
16. Morris, R. and J. Jackson, Absorption Towers, Butterworth Scientific Publications London, England (1953).
17. Bulletin S-29, Intalox Saddle Packing, The United States Stoneware Co. Akron, Ohio.
18. Bravo, J. L., "Effectively Fight Fouling of Packing," Chem. Eng. Prog., V. 89, No. 4, p. 72 (1993).
19. Kister, H. Z., Distillation Design, McGraw-Hill, Inc. New York, N.Y., (1992).
20. Schmidt, R. I., Chem. E. Symp. Fund., V. 6, p. 400 (1967).
21. Mass Transfer, Inc., "Absorption Distillation Optimization," p. 14 (1978).
22. Strigle, R. F., Jr. Packed Tower Design and Applications: Random and Structured Packings, 2nd Ed., Gulf Pub. Co., Houston, Texas (1994).
23. Kister, H. Z. and D. R. Gill, "Predict Flood Point and Pressure Drop for Modern Random Packings," Chem. Eng. Prog., V. 87, No. 2, p. 32 (1991).
24. Strigle, R. F. and F. Rukovena, "Packed Distillation Column Design," Chem. Eng. Prog. V. 75, p. 86, March (1979).
25. Zenz, F. A. Chem. Eng., p. 176, Aug. (1953).
26. Eckert, R. G., and Walter, Hydrocarbon Processing, 43 (2), p. 107 (1964).
27. Hsu, Shih-liang, "Packing Pressure Drop Estimated," Hydrocarbon Processing, V. 64, No. 7, p. 89 (1985).
28. Kaiser, V. "Correlate the Flooding of Packed Columns a New Way," Chem. Eng. Prog. V. 90, No. 6, p. 55 (1994).
29. Lobo, W. E., L. Friend, F. Hashmall, and F. Zenz, "Limiting Capacity of Dumped Tower Packings," Trans. AIChE, 41, 693 (1945).
30. Leva, M., Chem. Eng., Prog. Sym. Series, No. 10, 50, 151 (1954).
31. Leva, M. "Tower Packing and Packed Tower Design," 2nd. Ed. U.S., Stoneware Co., Akron, Ohio (1953).
32. Sakiadis, B. C., and A. I. Johnson, "Generalized Correlation of Flooding Rates," Ind. Eng. Chem., 46, 1229 (1954).
33. Robbin, L. A., "Improve Pressure Drop Prediction with a New Correlation," Chem. Eng., Prog., 87 May (1991).
34. "Report of Tests of No. 2.5 Nutter Ring™ at Reduced Loadings, released to Nutter Engineering" per Fractionation Research, Inc.
35. Nutter, D. E., F. C. Silvey, and B. K. Stober, "Random Packing Performance in Light Ends Distillation," Copyright by IChemE., undated, and publication not given.
36. Kunesh, John G. and A. Shariat, "Packing Efficiency Testing on a Commercial Scale with Good Reflux Distribution," Fractionation Research Inc., presented at AIChE Spring National Meeting, Houston, Texas, March (1993).
37. Bain, W. A. Jr. and O. A., Houben, Trans. AIChE., 40, p. 22 (1944).
38. Sherwood, T. K., G. H. Shipley and F. A. I., Holloway, "Flooding Velocities in Packed Columns," Ind. Eng. Chem., 30, p. 765 (1938).
39. Eckert, J. S., "Design Techniques for Sizing Packed Towers," Chem. Eng. Prog., 57, p. 54, Sept. (1961).
40. Blackwell, W. W., Chemical Process Design On A Programmable Calculator, McGraw-Hill Book Company, (1984).
41. Kessler, D. P. and P. C. Wankat, "Correlations for Column Parameters," Chem. Eng., p. 72, Sept. 26 (1988).
42. Nguyen, H. X., "Program Expedites Packed Tower Design," Chem. Eng., p. 181, Nov. 20 (1978).
43. Nguyen, H. X "Easy Way to Size Packed Towers," Hydro. Proc., p. 101, Feb. (1979).
44. Branan, C., "The Process Engineer's Pocket Handbook," Gulf Publishing Co., Houston, Texas, p. 87, (1976).
45. Eckert, J. S., "Selecting the Proper Distillation Column Packing," Chem. Eng. Prog. Vol. 66, No. 3, p. 39, March (1970).
46. Bolles, W., and J. F., Fair, "Improved Mass Transfer Model Enhances Packed-Column Design," Chem. Eng., 82, 14, p. 109, July (1982).
47. Norton Chemical Process Products, "Intalox^R High Performance Separation System," Bulletin IHP-1, Norton Chemical Process Products Corporation (1987).
48. Nutter Engineering Harsco Corp., "Nutter Ring™ Random Packing," Bulletin NR-2.
49. Zuiderweg, F. J and D. E. Nutter, "On the Evidence of Vapor Backmixing in Packed Columns in the Case of High-Pressure Distillation," Copyright by Institution of Chemical Engineers. Note: undated and publication not given.
50. Nutter, D. E., private communication.
51. Nutter, D. E., and D. Perry, "Packing Capacity/Pressure Drop Models," presented AIChE, New Orleans, La (1988).
52. Nutter, D. E., Silvery, F. C. and B. K., Stober, "Random Packing Performance in Lights Ends Distillation," IChemE. (U.K.), publication date not given.
53. Sherwood, T. K., and F. A. I., Holloway, "Performance of Packed Towers-Experimental Studies of Absorption and Desorption," Trans. AIChE, 36, 21 (1940).
54. Sherwood, T. K., and F. A. I., Holloway, "Performance of Packed Towers, Liquid Flow Data For Several Packings," Trans. AIChE, 36, 39 (1940).
55. Zenz, F. A., "What Every Engineer Should Know About Packed Tower Operations," Chem. Eng., p. 176, August (1953).

56. Ward, H. C. and J. T. Sommerfield, "New Flooding Equation," Hydrocarbon Processing, V. 61, p. 99, Oct. (1982).
57. Leva, M., "Flow Through Packings and Beds," Chem. Eng. 64, 261 (1957).
58. Tilson, Thesis, Mass Inst. Technology, Cambridge, Mass (1939).
59. Leva, M., Lucas, J. M., and H. H. Frahme, "Effect of Packing Supports on Mechanical Operation of Packed Towers," Ind. Eng. Chem., 46, No. 6, 1225 (1954).
60. Jesser, B. W., and J. C. Elgin, "Studies of Liquid Hold-up in Packed Towers," Trans. AIChE, 39, No. 3, 277 (1943).
61. Shulman, H. L., C. F. Ulrich, A. Z. and N. Wells, "Performance of Packed Columns, Total Static and Operating Holdups," AIChE. J., 1, No. 2, 247 (1955).
62. Shulman, H. L., C. F. Ulrich, A. Z., Wells, N., and A. Z. Proulx, "Holdup for Aqueous and Nonaqueous Systems," AIChE. J., Vol. 1, No. 2, 259 (1955).
63. Otake, T. and K. Okada, "Liquid Hold-up in Packed Towers, Operating and Holdup Without Gas Flow," Soc. Chem. Engrs. (Japan), 17, No. 7, 176 (1953).
64. Shulman, H. L., C. F. Ulrich, A. Z. Proulx and J. O. Zimmerman, "Wetted and Effective Interfacial Areas, Gas and Liquid phase Mass Transfer Rates," AIChE. J., No. 2, 253 (1955).
65. Shulman, H. L., and J. E. Margolis, "Performance of Packed Columns," AIChE. J., Vol. 3, 157 (1957).
66. "Koch Sulzer, Rectification Columns," Bulletin KS-1, Knoch Engineering Co. Inc., Witchta, Kansas.
67. "Koch Flexipac ® Structured Packing," Bulletin KFP-4, Koch Engineering Co., Inc.
68. Intalox ® High Performance Structured Packing," Bulletiin IS-IR; Norton Chemical Process Products Corporation
69. Martin, C. L., J. L. Bravo, and J. R. Fair, "Performance of Structured Packing in Distillation Service-Experimental and Modeling Results," presented at New Orleans, LA, AIChE. Meeting, Mar. 7 (1988).
70. "A Family of Gempack® Packing for High Efficiency Fractionation," Bulletin 5140 Glitsch, Inc. (1993).
71. "Koch Flexgrid Structural Packing," Bulletin KFG-2, Koch Engineering Co. Inc.
72. Fair, J. R., and J. L. Bravo, "Distillation Columns Containing Structured Packing," Chem. Eng. Prog., p. 19., Jan (1990).
73. Bravo, J. L., Rocha, J. A., and J. R., Fair, Hydro. Processing, V. 65, p. 45 Jan (1986).
74. Stichlmair, J., J. L. Bravo, and J. R. Fair, Gas Separation Purification, V. 3, p. 19 (1989).
75. Bravo, J. L., Rocha, J. A., and J. R., Fair, Hydro. Processing, V. 64, p. 91 Jan (1985).
76. Hatfield, J. A., "High Efficiency Tower Packings and Responsive Control Schemes," Chem. Processing, p. 130, Sept. (1988).
77. Bravo, J. L., Rocha, J. A., and J. R., Fair, "Pressure drop in Structured Packings," Hydro. Processing, V. 65, p. 45 Mar (1986).
78. Shah, G. C., "Effectively Troubleshoot Structural Packing Distillation Systems," Chem. Eng. Prog., V. 87, p. 49, April (1991).
79. Dean, J. A., Turner, H. M., and B. C. Price, "How Packing Works in Dehydration," Hydrocarbon Processing, V. 70, No. 4, p. 47 (1991).
80. Hausch, G. W, Quotson, P. K., and K. D. Seeger, "Structured Packing at High Pressure," Hydrocarbon Processing, V. 71, No. 4, p. 67 (1992).
81. Nutter Engineering Co. Harsco Corp., Bulletin B-1, "High Efficiency Structured Packings."
82. "Packings Growing Role in Distillation," Chem. Week, p. 18, June 13 (1984).
83. Hufton, J. R., Bravo, J. L, and J. R. Fair, "Scale-up of Laboratory Data for Distillation Columns Containing Corrugated Metal-type Structured Packing," Ind. and Eng. Chem. Res. American Chem. Soc., V. 27, No. 11, p. 2096 (1988).
84. Kister, H. Z., Scherffius, J. Afsbar, K, and E. Abkar, "Realistically Predict Capacity and Pressure Drop for Packed Columns," AIChE. CEP, p. 28, July (2007).
85. Colburn, A. P., "The Simplified Calculation of Diffusional Processes, General Considerations of Two Film Resistance," Trans. AIChE., 32, 211 (1939).
86. Colburn, A. P., "Simplified Calculation of Diffusional Processes," Ind. Eng. Chem., 33, 459 (1941).
87. Cornell, Knapp, and Fair, Chem. Eng. Prog., 56 (7), p. 68 (1960).
88. Treybal, R. E., Mass Transfer Operations, McGraw-Hill Book Co., New York, N.Y. (1955).
89. Coulson, J. M., and J. F. Richardson, Chemical Engineering, McGraw-Hill Book Co., New York, N.Y. Vol. II, 719 (1955).
90. Sherwood, T. K., and R. L., Pigford, Absorption and Extraction, 2nd Ed., McGraw-Hill Book Co., Inc., New York, N.Y., p. 161 (1952).
91. Harrison, M. E., "Consider Three-phase Distillation in Packed Columns," Chem. Eng. Prog., V. 86, No. 11, p. 80 (1990).
92. Ellis, Chem. Eng., News, 31 (44), p. 4613 (1953).
93. Cornell, *et al*. Chem. Eng. Prog., (58) (8), p. 68 (1960).
94. Fair, J., Chem. Eng. Prog., 76, (15), July (1969).
95. Silvey and Keller, Proc. Intern., Symp. Dist. (Brighton, U.K.) 1970.
96. Whitt, F. R., "A Correlation For Absorption Column Packings," British Chem. Eng., July, p. 395 (1959).
97. Hands, C. H. G., and F. R. Whitt, "Design of Packed Distillation Columns IV. An Empirical Method for the Estimation of Column Height Using the HETP Concept," J. Appl. Chem. Mar. 1, p. 135 (1951).

98. Teller, A. J., “Packing Performance Below and Above Flooding,” preprint copy prior to presentation at AIChE Meeting (1953)
99. Planovski, Khimicheskay, Prom., 45, No. 3, p. 16- 20.
100. Hands, C. H. G., and F. R. Whitt, “Design of Packed Distillation Columns II, Operating Vapor Rates for Packed Distillation Columns,” J. Appl. Chem. Jan. 1, p. 19 (1951).
101. Eckert, J. S., “A New Look at Distillation Tower Packings Comparative Performance, “ Chem. Eng. Prog. V. 59, No. 5, p. 76 (1963).
102. Eckert, J. S., “No Mystery in Packed Bed Design,” Oil & Gas J., Aug. 24 (1970).
103. Koshy, T. D. and K. Rukovena, “Reflux and Surface Tension Effects on Distillation,” Hydrocarbon Processing, V. 65, No. 5, p. 64 (1986).
104. Bolles, W. L., and J. R. Fair, “Improved Mass – Transfer Model Enhances Packed – Column Design,” Chem. Eng. V. 89, No. 14, p. 109 (1982).
105. Molstad, M. C., McKinney, J. F., and R. G. Abbey, “Performance of Drip-Point Grid Tower Packings,” Trans. AIChE., Vol. 39, No. 5, 605 (1943).
106. Vital, T. J., *et al*. Hydrocarbon Processing, V. 63, No. 12, p. 95 (1984).
107. Porter, K. E., and J. D. Jenkins, IChemE. Symposium Series No. 56, V. 3, p. 75 (1979).
108. Joerg Koch, and Glenn Shiveler, Design Principles for Liquid–Liquid Extraction, CEP, p. 22-30, November (2015).
109. Frank, T. C., *et al*. “Liquid – Liquid Extraction and Other Liquid–Liquid Operations and Equipment”, Section 15 in Green, D. W., and R. H. Perry, eds. “Perry’s Chemical Engineers’ Handbook,” 8th ed., McGraw-Hill, New York, NY (2008).
110. Rao, B. K. Bhaskara, Modern Petroleum Refining Processes, 6th ed., Oxford & IBH Publishing Co., Pvt. Ltd., (2014).
111. Hournac. R., Hydrocarbon processing, p. 207, January (1981).
112. Carter, R. C., Oil & Gas Journ., p. 137, March (1954).
113. Rao, B.K.B and T. S. Banerjee, 1st Law Symp. IIT, Bombay, Nov. 22, 23, (1978).
114. Henry Ruston, J., IEC., 27, p. 309, (1937).
115. Treybal, R. E., Liquid Extraction, McGraw-Hill, p. 60, (1951).
116. E. J. Henley and J. D. Seader, Equilibrium-Stage Separation Operations in Chemical Engineering. New York, John Wiley & Sons, (1981).
117. Kremser, A., Natl, Petroleum News 22 (21): 43–49, (1930).
118. Couper, James, R., Penny, W. Roy., Fair, James, R., and Stanley W. Walas., Chemical Process Equipment—Selection and Design, Revised 2nd ed., Elsevier, (2010).
119. Rukovena, F. and T. D. Koshy, “Packed Distillation Tower Hydraulic Design Method and Mechanical Considerations,” Ind. and Eng. Chem. Res., Vol. 32, No. 10 (1993), p. 2400.
120. Fellinger, L., Sc. D., Thesis, Mass Inst. Technol. (1941).
121. Fitz, C. W., Shariat, A., and L. Spielgel, “Performance of Structured Packing at High Pressure,” presented at AIChE Spring National Meeting, Houston, Texas, March (1955).
122. “Glitsch – Grid,” Bulletin 217 -5, 9-72, Glitsch, Inc.

Glossary of Petroleum and Petrochemical Technical Terminologies

Abatement: **1**. The act or process of reducing the intensity of pollution. **2**. The use of some method of abating pollution. **3**. Putting an end to an undesirable or unlawful condition affecting the wastewater collection system.

Abrasion (Mechanical): Wearing away by friction.

Abrasive: Particles propelled at a velocity sufficient to cause cleaning or wearing away of a surface.

Absolute Porosity: The percentage of the total bulk volume, which is pore spaces, voids, or fractures.

Absolute Pressure: **1**. The reading of gauge pressure plus the atmospheric pressure. **2**. Gauge pressure plus barometric or atmospheric pressure. Absolute pressure can be zero in a perfect vacuum. Units, psia, bara. e.g., psia = psig + 14.7, bara = barg + 1.013.

Absolute Temperature: Temperature measurement starting at absolute zero. e.g., $^{o}R = {}^{o}F + 460$, $K = {}^{o}C + 273.16$

Absolute Viscosity: The measure of a fluid's ability to resist flow without regard to its density. It is defined as a fluid's kinematic viscosity multiplied by its density.

Absorbent: The material that can selectively remove a target constituent from another compound by dissolving it.

Absorption: A variation of fractionation. In a distillation column, the stream to be separated is introduced in vapor form near the bottom. An absorption liquid called lean oil is introduced at the top. The lean oil properties are such that as the two pass each other, the lean oil will selectively absorb components of the stream to be separated and exit the bottom of the fractionator as rich oil. The rich oil is then easily separated into the extra and lean oil in conventional fractionation.

Absorption Gasoline: Gasoline is extracted from wet natural gas by putting the gas in contact with oil.

Absorption Oil (Facilities): The wash oil is used to remove heavier hydrocarbons from the gas stream.

Accident: An event or sequence of events or occurrences, natural or man-made, that results in undesirable consequences and requires an emergency response to protect life and property.

Accumulator: A vessel that receives and temporarily stores a liquid used in the feedstock or the processing of a feed stream in a gas plant or other processing facility.

Acentric Factor: A correlating factor that gives a measure of the deviation in the behavior of a substance to that for an idealized simple fluid. It is a constant for each component and has been correlated with the component vapor pressure.

Acid Gas: **1**. A gas that contains compounds such as CO_2, H_2S, or mercaptans (RSH, where $R = C_nH_{2n+1}$, n= 1, 2) that can form an acid in solution with water. **2**. Group of gases that are found in raw natural gas and are usually considered pollutants. Among these are CO_2, H_2S, and mercaptans. **3**. Any produced gas primarily H_2S and CO_2 that forms an acid when produced in water.

A. Kayode Coker. Petroleum Refining Design and Applications Handbook Volume 3, (961–1086) © 2022 Scrivener Publishing LLC

Acid Inhibitor: Acid corrosion inhibitor. It slows the acid attack on metal.

Acid Number: A measure of the amount of potassium hydroxide (KOH) needed to neutralize all or part of the acidity of a petroleum product. Also referred to as neutralization number (NN) or value (NV) and total acid number (TAN).

Acid-Soluble Oil (ASO): **1**. High boiling polymers are produced as an unwanted by-product in the alkylation processes. **2**. Polymers produced from side reactions in the alkylation process.

Acid Treating/Treatment: A process in which unfinished petroleum products, such as gasoline, naphthas, kerosene, diesel fuel, and lubricating oil stocks, are contacted with sulfuric acid to improve their color, odor, and other properties.

Acidity: The capacity of water or wastewater to neutralize bases. Acidity is expressed in milligrams per liter of equivalent calcium carbonate. Acidity is not the same as pH because water does not have to be strongly acidic (low pH) to have a high acidity. Acidity is a measure of how much base must be added to a liquid to raise the pH to 8.2.

AC Motor: Most of the pumps are driven by alternating current, three-phase motors. Such motors that drive pumps are usually fixed-speed drivers. DC motors are rarely used in process plants.

Actual Tray: A physical tray (contact device) in a distillation column, sometimes called a plate.

Activity of Catalyst: Activity generally means how well a catalyst performs with respect to reaction rate, temperature, or space velocity.

Adsorbents: Special materials like activated charcoal, alumina, or silica gel, used in an adsorption process that selectively cause some compounds, but not others to attach themselves mechanically as liquids.

Adsorption: **1**. A process for removing target constituents from a stream by having them condense on an adsorbent, which is then taken offline so the target constituents can be recovered. **2**. The process by which gaseous components adhere to solids because of their molecular attraction to the solid surface.

Alarms: Process parameters (levels, temperatures, pressures, flows) are automatically controlled within a permissible range. If the parameter moves outside this range, it sometimes activates both an audible and a visual alarm. If the panel board operator fails to take corrective action, a trip may also then be activated.

Alcohol: The family name of a group of organic chemical compounds composed of carbon, hydrogen, and oxygen. The series of molecules vary in chain length and are composed of a hydrocarbon plus a hydroxyl group, CH_3 (CH_2) n – OH (e.g., methanol, ethanol, tertiary butyl alcohol).

Alkanolamine: An organic nitrogen bearing compound related to ammonia having at least one, two, or three of its hydrogen atoms substituted with at least one, two, or three linear or branched alkanol groups where only one or two could also be substituted with a linear or branched alkyl group (i.e., methyl diethanolamine MDEA). The number of hydrogen atoms substituted by alkanol or alkyl groups at the amino site determines whether the alkanolamine is primary, secondary, or tertiary.

Alkylate: **1**. The gasoline produced by an alkylation process. It is made by combining the low boiling hydrocarbons catalytically to obtain a mixture of high-octane hydrocarbons boiling in the gasoline range. **2**. The product of an alkylation reaction. It usually refers to the high octane product from alkylation units. This alkylate is used in blending high octane gasoline.

Alkylation: **1**. A refining process for chemically combining isobutane (iC_4H_{10}) with olefin hydrocarbons [e.g., propylene (C_3H_6), butylenes (C_4H_8)] through the control of temperature and pressure in the presence of an acid catalyst. **2**. A refining process in which light olefins primarily a mixture of propylene (C_3H_6), butylenes (C_4H_8), and/or amylenes are combined with isobutane (iC_4H_{10}) over an acid catalyst to produce high octane gasoline (highly branched $C_5 – C_{12}$ i-paraffins), called alkylate. The commonly used catalysts are sulfuric acid (H_2SO_4) and hydrofluoric acid (HF). The major constituents of alkylate are isopentane and isooctane (2,2,4 – trimethyl pentane, TMP), the latter possessing an octane number of 100. The product, alkylate, is an isoparaffin, has high octane value, and is blended with motor and aviation gasoline to improve the antiknock value of the fuel.

Alkylate Bottoms: A thick, dark brown oil containing high molecular-weight polymerization products of alkylation reactions.

Aluminum Chloride Treating: A quality improvement process for steam cracked naphthas using aluminum chloride ($AlCl_3$) as a catalyst. The process improves the color and odor of the naphtha by the polymerization of undesirable olefins into resins. The process is also used when the production of resins is desirable.

American Petroleum Institute (API): An association, which among many things sets technical standards for measuring, testing, and other types of handling of petroleum.

Amine Treating: Contacting of a gas or light hydrocarbon liquid with an aqueous solution of an amine compound to remove the hydrogen sulfide (H_2S) and carbon dioxide (CO_2).

Anaerobic Digestion: This is a collection of processes by which microorganisms break down biodegradable material in the absence of oxygen. The process is used for industrial or domestic purposes to manage waste and/or to produce fuels. Much of the fermentation is used industrially to produce food and drink products, as well as home fermentation uses anaerobic digestion.

The digestion process begins with bacterial hydrolysis of the input materials. Insoluble organic polymers such as carbohydrates are broken down to soluble derivatives that become available for other bacteria. It is used as part of the process to treat biodegradable waste and sewage sludge. As part of an integrated waste management system, anaerobic digestion reduces the emission of landfill gas into the atmosphere. Anaerobic digestion is widely used as a source of renewable energy. The process produces biogas, consisting of methane, carbon dioxide, and traces of other "contaminant" gases. The biogas can be used directly as fuel in combined heat and power gas engines or upgraded to natural gas–quality biomethane. The nutrient-rich digestate also produced can be used as fertilizer.

Aniline Point: The minimum temperature for complete miscibility of equal volumes of aniline and the test sample. The test is considered an indication of the paraffinicity of the sample. The aniline point is used as a classification of the ignition quality of diesel fuels.

Antiknock Agent: **1**. Is a gasoline additive used to reduce engine knocking and increase the fuel's octane rating by raising the temperature and pressure at which auto-ignition occurs. The mixture is gasoline or petrol, when used in high compression internal combustion engines, has a tendency to knock (also referred to as pinging, or pinking) and/or to ignite early before the correct time spark occurs (pre-ignition, refers to engine knocking). **2**. The most wanted and widely used additives in gasoline are the antiknock compounds. They assist to enhance the octane number of gasoline. Lead in the form of tetraethyl lead (TEL) or tetramethyl lead (TML) is a good antiknock compound. TEL helps to increase the octane number of gasoline without affecting any other properties, including vapor pressure, but when used alone in gasoline gives rise to troublesome deposits.

Antiknock Index: The Research Octane Number (RON) test simulates driving under mild conditions while the Motor Octane Number (MON) test simulates driving under severe conditions, i.e., under load and at high speed.

The arithmetic average of RON and MON that gives an indication of the performance of the engine under the full range of conditions is referred to as Antiknock Index (AKI). It is determined by:

$$\text{Antiknock Index (AKI)} = \frac{\text{RON} + \text{MON}}{2}$$

Antiknock Quality (Octane Number): Knocking is a characteristic property of motor fuels that governs engine performance and is expressed in terms of octane number. It depends on the properties of hydrocarbon type and nature. Octane number is the percentage of isooctane in the reference fuel, which matches the knocking tendency of the fuel under test. Research octane number (RON) and motor octane number (MON) are two methods used and are measured with a standard single-cylinder, variable compression ratio engine. For both octane numbers, the same engine is used but operated at different conditions. The distinctions between two octane numbers (RON and MON) measurement procedures are engine speed, the temperature of admission, and spark advance. The motor method captures the gasoline at high engine speeds and loads, and the research octane method at a low speed depends on the fuel characteristics. The MON is normally 8–10 points lower than the RON. A high tendency to auto-ignite, or low octane rating, is undesirable in a gasoline engine, but desirable in a diesel engine.

$$\text{Antiknock index (AKI)} = (\text{RON} + \text{MON})/2.$$

API Gravity: A method for reporting the density of petroleum streams. It is defined as:

$^\circ\text{API} = \left[\frac{141.5}{\text{Sp.Gr} @ 60/60^\circ\text{F}} - 131.5\right]$, where Sp.Gr is the specific gravity relative to water. °API gravity is reported at a reference temperature of 60°F (15.9°C).

The scale allows representation of the gravity of oils, which on the specific gravity 60/60°F scale varies only over a range of 0.776 by a scale that ranges from less than 0 (heavy residual oil) to 340 (methane).

According to the expression, 10°API indicates a specific gravity of 1 (equivalent to water-specific gravity). Thus, higher values of API gravity indicate lower specific gravity and therefore lighter crude oils, or refinery products and vice versa. As far as crude oil is concerned, lighter API gravity value is desired as more amount of gas fraction, naphtha, and gas oils can be produced from the lighter crude oil than with the heavier crude oil. Therefore, crude oil with high values of API gravity is expensive to produce due to its quality.

Classification of crude oils

Crude category	**°API gravity**
Light crudes	°API > 38
Medium crudes	38 > °API > 29
Heavy crudes	29 > °API > 8.5
Very heavy crudes	°API < 8.5

The higher the API gravity, the lighter the compound. Light crudes generally exceed 38°API, and heavy crudes commonly are crudes with an °API of 22 or below. Intermediate crudes fall in the range of 22–38 °API. (See Figures 1a and 1b).

Aromatics: **1**. A group of hydrocarbons is characterized by having at least one benzene ring-type structure of six carbon atoms with three double and three single bonds connecting them somewhere in the molecule. The general formula is C_nH_{2n-6} where n = 6, 7, 8, etc. The simplest is benzene, plus toluene and the xylenes. Aromatics in gas oils and residues can have many, even scores of rings. **2**. The three aromatic compounds—benzene (C_6H_6), toluene (C_7H_8), and xylene (C_8H_{10}).

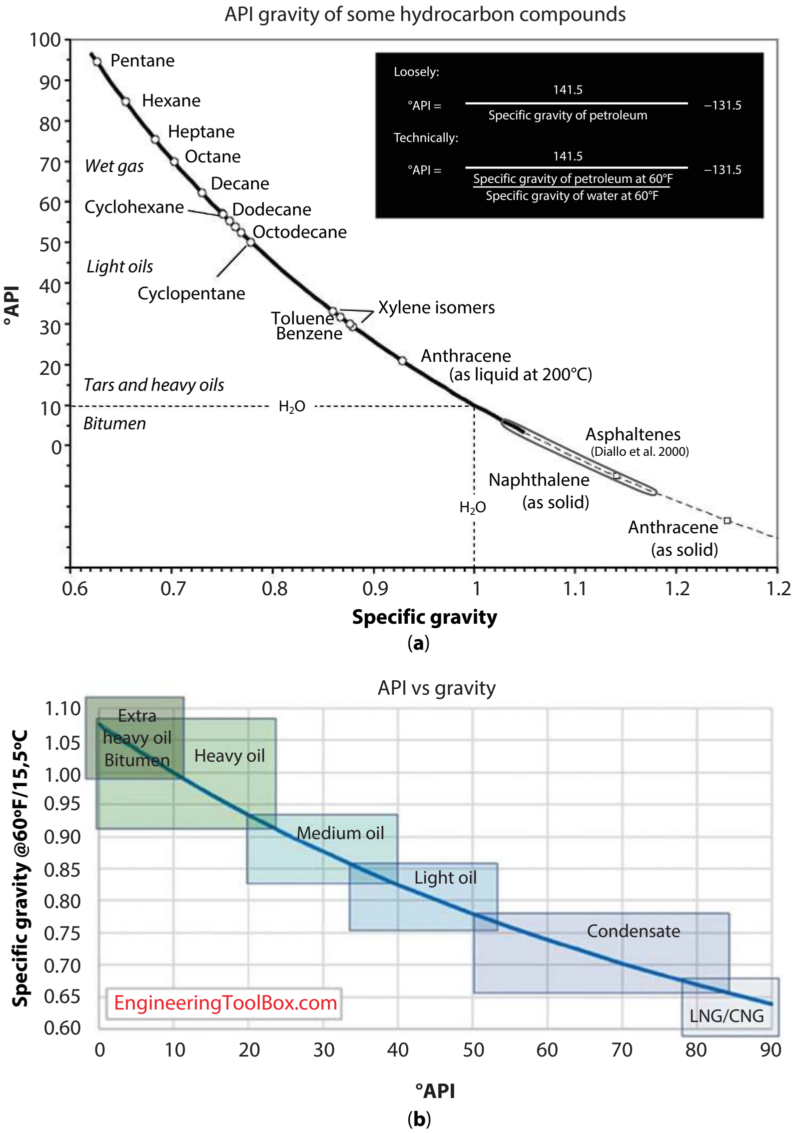


Figure 1 (a) A plot of °API vs. specific gravity of hydrocarbons compounds. (b) Specific gravity vs. °API of hydrocarbons compounds (Source: EngineeringToolBox.com)

As Low As Reasonably Practicable (ALARP): The principle that no industrial activity is entirely free from risk and that it is never possible to be sure that every eventuality has been covered by safety precautions, but that there would be a gross disproportion between the cost in (money, time, or trouble) of additional preventive or protective measures, and the reduction in risk in order to achieve such low risks (Figure 2).

Asphalt: **1**. A heavy semi-solid petroleum product that gradually softens when heated and is used for surface cementing. Typically, brown or black in color, it is composed of high carbon to hydrogen hydrocarbons. It occurs naturally in crude oil or can be distilled or extracted. **2**. The end product used for area surfacing consists of refinery asphalt mixed with aggregation. **3**. Heavy tar-like residue from distillation of some types of crude oil. Asphalt components are high molecular weight derivatives of aromatic compounds. Not all asphalt materials are suitable for use as building agents in road pavement.

Asphaltenes: Highly condensed masses of high molecular weight aromatic compounds. They exist in petroleum residuum as the center of colloidal particles or micelles. The asphaltenes are kept in solution by an outer ring of

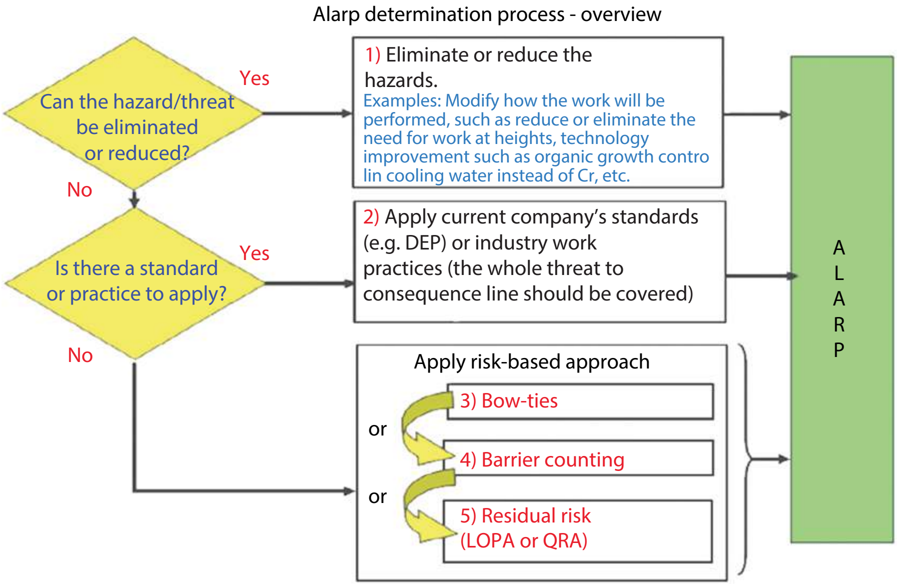


Figure 2 ALARP determination process overview. DEP = Design Engineering Practice.

aromatic compounds of lower molecular weight. They can precipitate when the continuous nature of the surrounding ring of aromatics is broken down by cracking processes.

Assay Data: Laboratory test data for a petroleum stream, including laboratory distillation, gravity, compositional breakdown, and other laboratory tests. Numerous important feed and product characterization properties in refinery engineering include:

1.	API gravity
2.	Watson Characterization factor
3.	Viscosity
4.	Sulfur content, wt %
5.	Nitrogen content, wt %
6.	Carbon residue, wt%
7.	Salt content
8.	Metal contents
9.	Asphaltene, %
10.	Naphthenes, %
11.	True boiling point (TBP) curve
12.	Pour point
13.	Cloud point
14.	Freeze point
15.	Aniline point

16.	Flash and fire point
17.	ASTM distillation curve
18.	Octane number
19.	Conradson carbon
21.	Reid vapor pressure
22.	Bottom sediment and water (BS &W)
23.	Light hydrocarbon yields (C_1 – C_5)

The crude quality is getting heavier worldwide. Existing refineries that are designed to handle normal crudes are being modified to handle heavy crude. New technology for upgrading is used to obtain clean and light products from lower cost feeds. The crude assay will determine the yields of different cuts and consequently the refinery configuration.

Associated Natural Gas: Natural gas that is dissolved in crude in the reservoir and is co-produced with the crude oil.

ASTM: American Society of Testing and Materials. Nearly all of the refinery product tests have been standardized by ASTM.

ASTM Distillation: A standardized laboratory batch distillation for naphthas and middle distillates carried out at atmospheric pressure without fractionation.

ASTM Distillation Range: Several distillation tests are commonly referred to as "ASTM distillations." These are usually used in product specifications. These ASTM distillations give results in terms of percentage distilled versus temperature for a sample laboratory distillation with no fractionation. The values do not correspond to those of refinery process distillations, where fractionation is significant.

ASTM D86 Distillation: Of an oil fraction takes place at laboratory room temperature and pressure. Note that the D86 distillation will end below an approximate temperature of 650°F (344°C), at which petroleum oils begin to crack at one atmospheric pressure.

ASTM D1160 Distillation: Of an oil fraction is applicable to high-boiling oil samples (e.g., heavy heating oil, cracker gas oil feed, residual oil, etc.) for which there is significant cracking at atmospheric pressures. The sample is distilled at reduced pressure, typically at 10 mm Hg, to inhibit cracking. In fact, at 10 mm Hg, we can distill an oil fraction up to temperatures of 950–1000°F (510–538°C), as reported on a 760 mm Hg basis. The reduced pressure used for D1160 distillation produces a separation of components that is more ideal than that for D86 distillation.

ASTM D2887 Distillation: Of oil fraction is a popular chromatographic procedure to "simulate" or predict the boiling point curve of an oil fraction. We determine the boiling point distribution by injecting the oil sample into a gas chromatograph that separates the hydrocarbons in a boiling-point order. We then relate the retention time inside the chromatograph to the boiling point through a calibration curve.

ASTM End Point of Distillates: End point is an important specification or way of describing gasolines, naphthas, or middle distillates. It is the approximate relationship between the end point of a fraction and its True Boiling Point (TBP) and other cut points.

Atmospheric distillation: **1**. The refining process of separating crude oil components at atmospheric pressure by heating to temperatures of 600–750°F (316–400°C) (depending on the nature of the crude oil and desired products) and subsequent condensing of the fractions by cooling. **2**. Distillation/Fractionation of crude oil into various cuts/fractions under atmospheric conditions. The more volatile components (i.e., lower boiling points) rise through trays/bubble caps and are condensed at various temperatures and the least volatile components, short and long residues (i.e., higher boiler points), are removed as bottom products.

Atmospheric Crude Oil Distillation: The refining process of separating crude oil components at atmospheric pressure by heating to temperatures of about 600–750°F (316–400°C) (depending on the nature of the crude oil and desired products) and subsequent condensing of the fractions by cooling.

Atmospheric Gas Oil (AGO): A diesel fuel and No. 2 heating oil blending stock obtained from the crude oil as a side stream from the atmospheric distillation tower.

Atmospheric Reduced Crude (ARC): The bottoms stream from the atmospheric distillation tower.

Atmospheric Residuum: The heaviest material from the distillation of crude oil in a crude distillation column operating at a positive pressure.

Autoignition: The spontaneous ignition and resulting in a rapid reaction of a portion of or all the fuel–air mixture in the combustion chamber of an internal combustion engine. The flame speed is many times greater than that following normal ignition.

Autoignition Temperature (AIT): **1**. The lowest temperature at which a gas will ignite after an extended time of exposure. **2**. The lowest temperature at which a flammable gas or vapor air mixture will ignite from its own heat source or a contacted heat source without the necessity of a spark or a flame.

Aviation Gasoline Blending Components: Naphthas that will be used for blending or compounding into finished aviation gasoline (e.g., straight-run gasoline, alkylate, reformate, benzene, toluene, xylenes). Excludes oxygenates (alcohols, ethers), butanes, and pentanes. Oxygenates are reported as other hydrocarbons, hydrogen, and oxygenates.

Aviation Gasoline (Finished): A complex mixture of relatively volatile hydrocarbons with or without small quantities of additives, blended to form a fuel suitable for use in aviation reciprocating engines. Fuel specifications are provided in ASTM Specification D910 and Military Specification MIL-G-5572. Note: Data on blending components are not counted in data on finished aviation gasoline.

Azeotrope: A constant boiling point mixture for which the vapor and liquid have identical composition. Azeotropes cannot be separated from conventional distillation.

Backflow: **1**. A flow condition, caused by differential pressure, resulting in the flow of liquid into the potable water supply system from sources other than those intended; or the backing up of liquid, through a conduit or channel, in a direction opposite to normal flow. **2**. Return flow from an injection of a fluid into a formation.

Back Pressure: A pressure caused by a restriction or fluid head that exerts an opposing pressure to flow.

Barrel: A volumetric measure of refinery feedstocks and products equal to 42 US gal.

Barrels Per Calendar Day (BPCD or B/CD): Average flow rates based on operating 365 days per year. The amount of input that a distillation facility can process under usual operating conditions. The amount is expressed in terms of capacity during a 24-hour period and reduces the maximum process capability of all units at the facility under continuous operation to account for the following limitations that may delay, interrupt, or slow down production: The capability of downstream facilities to absorb the output of crude oil processing facilities of a given refinery. No reduction is made when a planned distribution of intermediate streams through other than downstream facilities are part of a refinery's normal operation; the types and grades of inputs to be processed; the types and grades of products expected to be manufactured; the environmental constraints associated with refinery operations; the reduction of capacity for scheduled downtime due to such conditions as routine inspection, maintenance, repairs, and turnaround, and the reduction of capacity for unscheduled downtime due to such conditions.

Barrels Per Stream Day (BPSD or B/SD): The maximum number of barrels of input that a distillation facility can process within a 24-hour period when running at full capacity under optimal crude and product slate conditions with no allowance for downtime. This notation equals barrels per calendar day divided by the service factor.

Basic Process Control System (BPCS): A system that responds to input signals from the process, its associated equipment, other programmable systems, and/or an operator and generates output signals causing the process and its associated equipment to operate in the desired manner but which does not perform any safety instrumented functions (SIF) with a claimed Safety Instrumented Level, SIL ≥ 1.

Battery Limits (BL): The periphery of the area surrounding any process unit, which includes the equipment for the particular process.

Baume gravity: Specific gravity of liquids expressed as degrees on the Baume scale. For liquids lighter than water,

$$\mathrm{Sp.Gr}\,@15.6\Big/15.6^\circ\mathrm{C} = \frac{140}{130 + \deg\mathrm{Be}}$$

For liquids heavier than water

$$\mathrm{Sp.Gr}\,@15.6\Big/15.6^\circ\mathrm{C} = \frac{145}{145 - \deg\mathrm{Be}}$$

Bbl: Abbreviation for a quantity of 42 US gal.

Benchmark crude: A reference crude oil with which the prices of other crudes are compared.

Benzene (C_6H_6): An aromatic hydrocarbon present in small proportion in some crude oils and made commercially from petroleum by the catalytic reforming of naphthenes in petroleum naphtha. It is also made from coal in the manufacture of coke. Used as a solvent, in manufacturing detergents, synthetic fibers, and petrochemicals and as a component of high-octane gasoline.

Bernoulli equation: A theorem in which the sum of the pressure-volume, potential, and kinetic energies of an incompressible and non-viscous fluid flowing in a pipe with the steady flow with no work or heat transfer is the same anywhere within a system. When expressed in head form, the total head is the sum of the pressure, velocity, and static head. It is applicable only for incompressible and non-viscous fluids as:

SI Units

$$\frac{P_1}{\rho_1} + \frac{v_1^2}{2g} + z_1 = \frac{P_2}{\rho_2} + \frac{v_2^2}{2g} + z_2 + h_f$$

where h_f is the pipe friction from point 1 to point 2 may be referred to as the head loss in meters of fluid.

Imperial Units

$$\frac{144P_1}{\rho_1}+\frac{v_1^2}{2g_c}+z_1\frac{g}{g_c}=\frac{144P_2}{\rho_2}+\frac{v_2^2}{2g_c}+z_2\frac{g}{g_c}+h_f$$

where h_f is the pipe friction from point 1 to point 2 in foot-pounds force per pound of flowing fluid; this is sometimes referred to as the head loss in feet of fluid.

where, P is pressure, ρ is density, g_c is a conversion factor $\left(32.174\frac{lb_m}{lb_f}\cdot\frac{ft}{s^2}\right)$, g is the acceleration due to gravity (32 ft/s^2), v is velocity, z is elevation, and h_f is frictional head loss. It is a statement of the law of the conservation of energy, which was formulated by Daniel Bernoulli in 1738 (Figure 3).

Bioenergy with carbon capture and Storage (BECCS): **1.** Is the process of extracting bioenergy from biomass and capturing and storing the carbon. **2.** Is the process during which carbon is captured and stored. If carefully managed, using sustainable biomass, BECCS can generate "negative emissions" because while providing energy it also captures and stores the atmospheric carbon monoxide (CO) that is absorbed by plants as they grow.

Bitumen: That portion of petroleum, asphalt, and tar products that will dissolve completely in carbon disulfide (CS_2). This property permits a complete separation from foreign products not soluble in carbon disulfide.

Blast: A transient change in gas density, pressure (either positive or negative), and velocity of the air surrounding an explosion point.

Blending: One of the final operations in refining, in which two or more different components are mixed together to obtain the desired range of properties in the final product.

Blending Components: Modern gasoline is a blend of various refinery streams produced by distillation, cracking, reforming, and polymerization together with additives to achieve the specific fuel performance requirements.

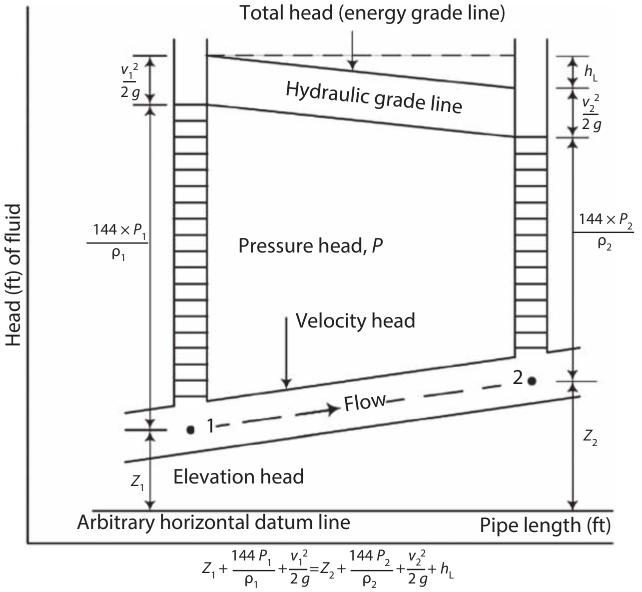


Figure 3 Distribution of fluid energy in a pipeline.

Blending Octane Number: When blended into gasoline in relatively small quantities, high-octane materials behave as though they had an octane number higher than shown by laboratory tests on the pure material. The effective octane number of the material in the blend is known as the blending octane number.

Blending Plant: A facility that has no refining capability but is either capable of producing finished motor gasoline through mechanical blending or blends oxygenates with motor gasoline.

Blending Value (Hydrocarbon): In octane ratings of a hydrocarbon made on blends of 20% hydrocarbon plus 80% of a 60:40 mixture of isooctane (iC_8H_{18}) and n-heptane (nC_7H_{16}), the blending octane number is a hypothetical value obtained by extrapolation of a rating of 100% concentration of the hydrocarbon.

Blocked Operation: A set of operating conditions and procedures that apply to particular feedstock and/or set of product specifications for a process.

Boiler: **1**. A closed vessel in which a liquid is heated or heated and evaporated. Boilers are often classified as steam or hot water, low pressure or high pressure, and capable of burning one fuel or a number of fuels. **2**. Vessel in which a liquid is heated with or without vaporization; boiling need not occur.

Boiler Feed Pump: A pump that returns condensed steam, makeup water, or both directly to the boiler.

Boiler Feed Water: Water supplied to a boiler by pumping.

Boiling Liquid Expanding Vapor Explosion (BLEVE): 1. The nearly instantaneous vaporization and corresponding release of energy of a liquid upon its sudden release from a containment under pressure than atmospheric pressure and at a temperature above its atmospheric boiling point. **2**. A type of rapid phase transition in which a liquid contained above its atmospheric boiling point is rapidly depressurized, causing a nearly instantaneous transition from liquid to vapor with a corresponding energy release. A **BLEVE** is often accompanied by a large fireball if a flammable liquid is involved, since an external fire impinging on the vapor space of a pressure vessel is a common **BLEVE** scenario. However, it is not necessary for the liquid to be flammable to have a **BLEVE** to occur.

Blowdown: The disposal of voluntary discharges of liquids or condensable vapors from process and vessel drain valves, thermal relief or pressure relief valves.

Blowout: An uncontrolled flow of gas, oil, or other well fluids from a wellbore at the wellhead or into a ground formation, caused by the formation pressure exceeding the drilling fluid pressure. It usually occurs during drilling on unknown (exploratory) reservoirs.

Boiling Point: **1**. Heat a liquid and its vapor pressure increases. When the liquid's vapor pressure equals the pressure in the vessel, the liquid starts to boil. The temperature at which this boiling starts is the liquid's boiling temperature. **2**. Typically refers to the temperature at which a component or mixture of components starts to vaporize at a given pressure. When used in petroleum refining, it is usually synonymous with the normal boiling point (i.e., boiling point at one atmosphere). **3**. The temperature at which the pressure exerted by molecules leaving a liquid equals the pressure exerted by the molecules in the air above it. A free-or-all of the molecules leaving the liquid then ensures. In a solution, the boiling point will be increased by a number that depends on the number of particles in the solution:

$$\text{delta (T)} = K_b \text{ x (number of solute molecules per liter)}$$

where

delta (T) = the rise in the boiling point.

K_b = the ebulllioscopic constant and varies from one solvent to another.

Boiling Range: **1**. The spread of temperatures over which oil starts to boil or distill vapors and proceeds to complete evaporation. The boiling range is determined by ASTM test procedures for specific petroleum products. It is measured in °F or (°C). **2**. The lowest through to highest boiling temperatures for a petroleum stream when distilled. Boiling ranges are often reported on a TBP (true boiling point) basis, i.e., as normal boiling points.

Boiling Temperature: The temperature at which steam bubbles begin to appear within a liquid. When the fluid is a pure compound, the boiling point is unique for each pressure.

Boil Off: A small amount of LNG evaporates from the tank during storage, cooling the tank and keeping the pressure inside the tank constant and the LNG at its boiling point. A rise in temperature is encountered by the LNG being vented from the storage tank.

Boil Off Vapor: Usually refers to the gases generated during the storage or volatile liquefied gases such as LNG. Natural gas boils at slightly above -261°F (-163°C) at atmospheric pressure and is loaded, transported, and discharged at this temperature, which requires special materials, insulation, and handling equipment to deal with the low temperature and the boil-off vapor.

Boot, Boot Cooler: The section of a distillation column below the trays. For columns with very hot feeds, a portion of the bottom product is cooled and circulated through the boot or lowers the temperature of the liquid in the boot and prevents the depositing of coke. Many vacuum distillation columns have boot coolers.

Bottoms: **1**. The heavy fractions or portions of crude oil that do not vaporize during fractionation/distillation. **2**. The accumulation of sediments, mud, and water in the bottoms of lease tanks. **3**. The product coming from the bottom of a fractionating column. In general, the higher-boiling residue is removed from the bottom of a fractionating tower. **4.** The liquid level is left in a tank after it has been pumped "empty" and the pump loses suction.

Bowtie analysis: **1**. A qualitative risk analysis that portrays events and consequences on either side of a "bowtie." Barriers or safeguards are shown in between the two sides. It depicts the risks in ways that are readily understandable to all levels of operations and management. **2**. A type of qualitative safety review where cause scenarios are identified and depicted on the pre-event side (left side) of a bow-tie diagram. Credible consequences and scenarios outcomes are depicted on the post-event side (right side) of the diagram, and associated barrier safeguards are included (Figure 4).

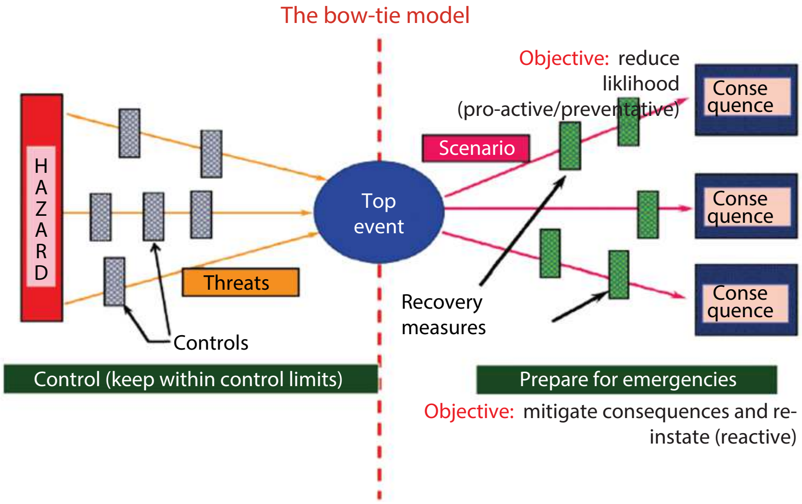


Figure 4 The bow-tie analysis.

Brackish Water: Indefinite term meaning water with small amounts of salt. Saltier than freshwater.

Brainstorming: A group problem-solving technique that involves the spontaneous contribution of ideas from all members of the group primarily based on their knowledge and experience.

Brent: A large oil field in the U.K. sector of the North Sea. Its name is used for a blend of crudes widely used as a price marker or benchmark for the international oil industry. Brent crude currently has an average quality of 38°API.

Brent Blend: A light sweet crude oil produced in the North Sea; a benchmark for pricing of many foreign crude oils.

Bright Stock: Heavy lube oils (frequently the vacuum still bottoms) from which asphaltic compounds, aromatics, and waxy paraffins have been removed. Bright stock is one of the feeds to a lube oil blending plant.

British thermal unit (Btu): A standard measure of energy; the quantity of heat required to raise the temperature of 1 pound of water by 1°F.

Bromine Index: Measure of the amount of bromine reactive material in a sample; ASTM D-2710.

Bromine Number: A test that indicates the degree of unsaturation in the sample (olefins and diolefins); ASTM D-1159.

BTX: The acronyms for the commercial petroleum aromatics benzene, toluene, and xylene.

Bubble Cap: **1**. It is an inverted cup with a notched or slotted periphery to disperse the vapor in small bubbles beneath the surface of the liquid on the bubble plate in a distillation column. The bubble caps cause the vapor coming from the bottom to come in intimate contact with the liquid sitting on the tray. **2**. A bubble cap tray has a riser or chimney fitted over each hole and a cap that covers the riser. The cap is mounted so that there is a space between the riser and cap to allow the passage of vapor. Vapor rises through the chimney and is directed downward by the cap, finally discharging through the slots in the cap, and finally bubbling through the liquid on the tray (Figure 5).

Bubble Point: **1**. This is the same as the boiling point. When a liquid is at its bubble point, it is said to be saturated liquid at temperature and pressure. If we raise the pressure, the liquid's bubble point temperature goes up. *2*. The temperature and pressure at which a liquid first begins to vaporize into a gas. **3**. The temperature at which the first bubbles appear when a liquid mixture is heated. **4**. The temperature at which a component or mixture of components begins to vaporize at a given pressure. It corresponds to the point of 0% vaporization or 100% condensation. The pressure should be specified, if not one atmosphere. **5**. The pressure at which gas begins to break out of undersaturated oil and form a free gas phase in the matrix or a gas cap.

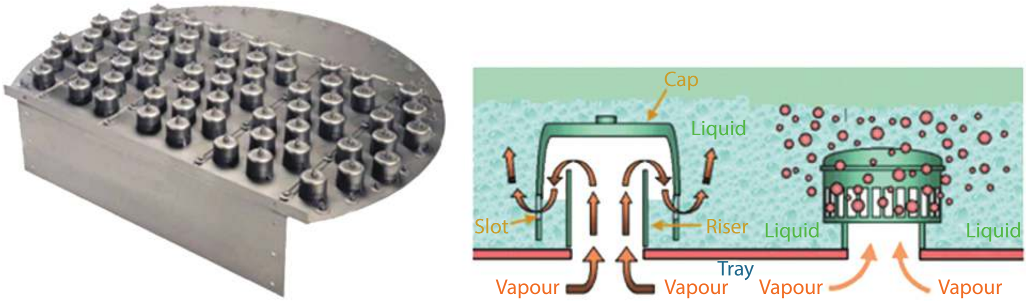


Figure 5 A bubble cap tray.

Bubble Tower or Column: A fractionating tower constructed in such a way that the vapors rising up pass through different layers of condensate on a series of plates. The less volatile portions of vapor condense in bubbling through the liquid on the plate, overflow to the next lower plate, and finally back to the boiler.

Bubble Tray: A horizontal tray fitted in the interior of a fractionating tower; meant to give intimate contact between rising vapors and falling liquid in the tower.

Bulk Properties: Provide a quick understanding of the type of the oil sample such as sweet or sour, light and heavy, etc. However, refineries require fractional properties of the oil sample that reflects the property and composition for specific boiling-point range to properly refine it into different end products such as gasoline, diesel, and raw materials for chemical process. Fractional properties usually contain paraffins, naphthenes, and aromatics (PNA) contents, sulfur content, nitrogen content for each boiling-point range, the octane number of gasoline, freezing point, cetane index, and smoke point for kerosene and diesel fuels.

Bulk Station: A facility used primarily for the storage and/or marketing of petroleum products, which has a total bulk storage capacity of less than 50,000 barrels and receives its petroleum products by tank car or truck.

Bulk Terminal: A facility used primarily for the storage and/or marketing of petroleum products, which has a total bulk storage capacity of 50,000 barrels or more and/or receives petroleum products by tanker, barge, or pipeline.

Bunker Fuel Oil: A heavy residual fuel oil used by ships, industry, and large-scale heating installations.

Butadiene (C_4H_6): A diolefin with two double bonds and two isomers. A colorless gas resulting from cracking processes. Traces result from the cat. cracking from catalytic dehydrogenation of butane (C_4H_{10}) or butylenes (C_4H_8) and in ethylene plants using butane, naphtha, or gas oil as feeds. Butadiene is principally used to make polymers like synthetic rubber and acrylonitrile butadiene styrene (ABS) plastics.

Butane (C_4H_{10}): A normally gaseous four-carbon straight-chain or branched-chain hydrocarbon extracted from natural gas or refinery gas streams. It includes normal butanes and refinery grade butanes and is designated in ASTM Specification D1835 and Gas Processors Association Specifications for commercial butane. Commercial butane is typically a mixture of normal and isobutene, predominantly normal. Hydrocarbons in the paraffin series with a general formula C_nH_{2n+2}, where n = 1, 2, 3, 4, 5, etc. To keep the liquid and economically stored, butane must be maintained under pressure or at low temperatures.

Butylene/Butene (C_4H_8): Hydrocarbons with several different isomers in the olefin series with a general formula C_nH_{2n}. Used in refining in an alkylation plant or in petrochemicals to make solvents and some polymers.

Carbon-to-Hydrogen Ratio: The carbon-to-hydrogen ratio is determined by the following:

$$\frac{C}{H} = \frac{74 + 15d}{26 - 15d}$$

where d is the specific gravity at 15°C

The carbon–hydrogen ratios of different products are:

LPG (d= 0.56)	= 4.68
Naphtha (d = 0.72)	= 5.57

Gasoline (d = 0.73)	= 5.64
ATF (d = 0.79)	= 6.067
SK (d = 0.795)	= 6.10
JP5 (d = 0.80)	= 6.14
HSD (d = 0.845)	= 6.50
LDO (d = 0.87)	= 6.72
LSHS (d = 0.98)	= 7.85
FO (d = 0.99)	=7.97

Calorific Value: **1**. A measure of the amount of energy released as heat when a fuel is burned. **2**. The quantity of heat produced by the complete combustion of a fuel. This can be measured dry or saturated with water vapor, net or gross.

It is a measure of the heat-producing capacity of the fuel. It is determined by:

$$Q_v = 12400 - 2100d^2$$

where

Q_v = calorific value, gross cals/g
d = density at 15°C

Note: 1 cal = 4.184 Joules

Calorific value (average) of different fuels

Fuel	**Calorific value, kcal/kg**
Naphtha	11330
Kerosene	11070
HSD	10860
Fuel Oil	10219
Charcoal	6900
Hard coal	5000
Fire wood	4750
Lignite-Brown coal	2310

Carbon Number: The number of carbon atoms in one molecule of a given hydrocarbon.

Carbon Dioxide Equivalent (CO_{2eq}): US Environmental Protection Agency (EPA) defines carbon dioxide equivalent as the number of metric tons of CO_2 emissions with the same global warming potential as one metric ton of another greenhouse gas, and is calculated using Equation A-1 in 40 CFR Part 98.

Carbon Residue: Carbon residue is a measure of the coe-forming tendencies of oil. It is determined by destructive distillation in the absence of air of the sample to a coke residue. The coke residue is expressed as the weight percentage of the original sample. There are two standard ASTM tests, Conradson carbon residue (CCR) and Ramsbottom carbon residue (RCR).

Carbon capture and storage (CCS): CCS or carbon capture and sequestration is the process of capturing carbon dioxide (CO_2) before it enters the atmosphere**.**

Carbon intensity: The amount of CO_2 emitted when generating a unit of electricity, measured in gram of CO_2 per kWh of electricity produced.

Carbon Capture Utilization and Storage (CCUS): The process of capturing carbon dioxide from industrial processes, power generation, certain hydrogen production methods, and greenhouse gas removal technologies such as bioenergy with carbon capture and storage and direct air capture. The captured CO_2 is then either used, for example in chemical processes, or stored permanently in disused oil and gas fields or naturally occurring geological storage sites.

Carbon Footprint: While carbon footprints are usually reported in tons of emissions (CO_2-equivalent) per year, ecological footprints are usually reported in comparison to what the planet can renew. This assesses the number of "earths" that would be required if everyone on the planet consumed resources at the same level as the person calculating their ecological footprint. The carbon footprint is one part of the ecological footprint. Carbon footprints are more focused than ecological footprints since they measure merely the emissions of gases that cause climate change into the atmosphere.

Carbon Leakage: **1.** Carbon leakage is the increase in CO_2 emissions outside the countries taking domestic mitigation action divided by the reduction in the emissions of these countries. It is expressed as a percentage and can be greater or less than 100%. **2.** Occurs when there is an increase in greenhouse gas emissions in one country as a result of emissions reduction by a second country with a strict climate policy.

Carbon leakage may occur for a number of reasons:

- If the emissions policy of a country raises local costs, then another country with a more relaxed policy may have a trading advantage. If demand for these goods remains the same, production may move offshore to the cheaper country with lower standards, and global emissions will not be reduced.
- If environmental policies in one country add a premium to certain fuels or commodities, then the demand may decline and their price may fall. Countries that do not place a premium on those items may then take up the demand and use the same supply, negating any benefit.

Carbon Price: A cost applied to carbon pollution to encourage polluters to reduce the amount of greenhouse gases they emit into the atmosphere.

Carbon Tax: 1. A carbon tax is a form of pollution tax. Its aim is to allow market forces to determine the most efficient way to reduce pollution**. 2.** It is an indirect tax—i.e., a tax on a transaction as opposed to direct tax, which taxes income. Carbon taxes are price instruments since they set a price rather than an emission limit. In addition to creating incentives for energy conservation, a carbon tax puts renewable energy such as wind, solar, and geothermal on a more competitive footing. Although carbon tax covers only CO_2 emissions, it can also cover other greenhouse gases, such as methane (CH_4) or nitrous oxide (NO_x), by calculating their global warming potential (GWP) relative to CO_2. When a hydrocarbon fuel such as coal, petroleum, or natural gas is burnt, much of its carbon is converted to CO_2. Greenhouse gas emissions cause climate change, which damages the environment and human health. **3**. A carbon tax puts renewable energy such as wind, solar, and geothermal on a more competitive footing.

Catalyst: A substance present in a chemical reaction that will promote, accelerate, or selectively direct a reaction, but does not take part in it by changing chemically itself. Sometimes a catalyst is used to lower the temperature or pressure at which the reaction takes place.

Catalyst-to-Oil Ratio (C/O): The weight of circulating catalyst fed to the reactor of a fluid-bed catalytic cracking unit divided by the weight of hydrocarbons charged during the same interval.

Catalytic Cracking: **1**. The refining process of breaking down the larger, heavier, and more complex hydrocarbon molecules into simpler and lighter molecules. Catalytic cracking is accomplished by the use of a catalyst and is an effective process for increasing the yield of gasoline from crude oil. Catalytic cracking processes fresh feeds and recycled feeds. **2**. A central process in reforming in which heavy gas oil range feeds are subjected to heat in the presence of a catalyst and large molecules crack into smaller molecules in the gasoline, diesel, and surrounding ranges. **3**. A petroleum refining process in which heavy hydrocarbon molecules are broken down (cracked) into lighter molecules by passing them over a suitable catalyst (generally heated). **4**. A method of cracking that uses a catalyst to convert hydrocarbons to positively charged carbonations, which then break down into smaller molecules. This can be carried out at much lower temperatures than thermal cracking—still hot 932–1112°F (500–600°C) as compared to around 1292°F (700°C). But that difference adds up to a lot of dollars.

Catalytically Cracked Distillates: These are obtained when high boiling non-gasoline hydrocarbons are heated under pressure in the presence of a catalyst to obtain lower boiling gasoline components. Catalytically cracked distillates usually have high octane numbers than straight-run gasoline.

Catalytic Cycle Stock: That portion of a catalytic cracker reactor effluent that is not converted to naphtha and lighter products. This material, generally 340°F (170°C), either may be completely or partially recycled. In the latter case, the remainder will be blended to products or processed further.

Catalytic Hydrocracking: A refining process that uses hydrogen and catalysts with relatively low temperatures and high pressures for converting middle boiling or residual material to high-octane gasoline, reformer charge stock., jet fuel, and/or high-grade fuel oil. The process uses one or more catalysts, depending upon product output, and can handle high sulfur feedstocks without prior desulfurization.

Catalytic Hydrotreating: A refining process for treating petroleum fractions from atmospheric or vacuum distillation units (e.g., naphthas, middle distillates, reformer feeds, residual fuel oils, and heavy gas oil) and other petroleum (e.g., cat cracked naphtha, coker naphtha, gas oil, etc.) in the presence of catalysts and substantial quantities of hydrogen. Hydrotreating includes desulfurization, removal of substances (e.g., nitrogen compounds) that deactivate catalysts, conversion of olefins to paraffins to reduce gum formation in gasoline, and other processes to upgrade the quantity of the fractions.

Catalytic Polymerization (cat. poly): A process in which propylene and/or butylenes components are chemically joined to produce gasoline. A phosphoric acid (HPO_3) catalyst is usually employed in the process.

Catalyst Promoter: A substance added to a catalyst to increase the fraction of the total catalyst area, which is useful for a reaction.

Catalytic Reforming: **1**. A refining process using controlled heat and pressure with catalysts to rearrange certain hydrocarbon molecules, thereby converting paraffinic and naphthenic hydrocarbons (e.g., low-octane gasoline boiling range fractions) into petrochemical feedstocks and higher octane stocks suitable for blending into finished gasoline. **2**. A process where low octane straight-run naphthas are chemically changed into high octane gasoline, called reformate, and to produce aromatics (BTX: benzene, toluene, and xylene) for petrochemical plants over a platinum (Pt) catalyst. The reformate has higher aromatic and cyclic hydrocarbon contents. Catalytic reforming is reported into two categories, namely:

Low Pressure. A processing unit operating at less than 225 psig measured at the outlet separator.

High Pressure: A processing unit operating at either equal to or greater than 225 psig measured at the outlet separator.

Catalyst Selectivity: The relative activity of a catalyst with respect to a particular component or compound in a mixture.

Catalyst Stripping: The introduction of steam at a point where spent catalyst leaves the reactor, in order to remove or strip the hydrocarbons retained on the catalyst.

Catastrophic Incident: An incident involving a major uncontrolled emission, fire, or explosion with an outcome effect zone that extends offsite into the surrounding community.

Cause: The reasons why deviation might occur.

Caustic Soda: Name used for sodium hydroxide (NaOH); used in refineries to treat acidic hydrocarbon streams to neutralize them.

Cavitation: **1**. The creating of high-speed, very low-pressure vapor bubbles that quickly and violently collapse. It is very detrimental to surfaces in the near proximity, and is often seen in a severe turbulent flow. **2**. Occurs during vaporization of a pumped fluid resulting in vibration, noise, and destruction of equipment. This is when the absolute pressure of the system equals the vapor pressure of the pumped fluid. In a centrifugal pump, it results in the damage of the impeller. **3**. When the pressure of liquid flowing into a centrifugal pump gets too low, the liquid boils inside the pump case and generates bubbles. The discharge pressure and flow become erratically low.

Centipoise (cP): A measure of viscosity related to centistokes by adjusting for density. **1**. Viscosity measurement, 1/1000th of a poise. **2.** A centipoise (cP) is 1/1000th of a poise (P), which is the fundamental unit of dynamic viscosity in the CGS system of units. In the SI system of units, the fundamental unit of dynamic viscosity is the Pascal second (Pa.s) and is equivalent to 10P.

Centistoke (cSt): Is 1/100th of a Stoke (St), which is the fundamental unit of kinematic viscosity in the CSG system of units. In the SI system of units, the fundamental unit of kinematic viscosity is the millimeter squared per second (mm^2/s), which is equivalent to the centistokes.

Cetane (Hexadecane, $C_{16}H_{34}$): An alkane hydrocarbon with a chemical formula $C_{16}H_{34}$ used as a solvent and in cetane number determinations. **1**. A pure paraffin hydrocarbon used as standard reference fuel in determining the ignition qualities of diesel fuels. **2**. A number calculated from the API gravity and the D86 50% distilled for a petroleum stock. It is used to rate the performance of a fuel in diesel engines. It is arbitrarily given a cetane number of 100.

Cetane Index: **1**. A number calculated from the average boiling point and gravity of a petroleum fraction in the diesel fuel boiling range, which estimates the cetane number of the fraction according to ASTM D976. An indication of the carbon–hydrogen ratio. **2.** An empirical method for determining the cetane number of diesel fuel by a formula based on API gravity and the mid-boiling point (ASTM D975) (see, for example, http://www.epa.gov/nvfel/testproc/121.pdf).

Cetane Number: **1**. The percentage of pure cetane in a blend of cetane and alpha- methyl-naphthalene matches the ignition quality of a diesel fuel sample. This quality, specified for middle distillate fuel, is the opposite of the octane number of gasoline. It is an indication of ease of self-ignition. **2**. A term for expressing the ignition quality of diesel fuel. **3**. A measure of the ignition quality of diesel fuel, expressed as a percentage of cetane that must be mixed with methyl naphthalene to produce the same ignition performance as the diesel fuel being rated. The higher the number, the more easily the fuel is ignited under compression. It is an important factor in determining the quality of diesel fuel. In short, the higher the cetane number, the more easily the fuel will combust in a compression setting (such as a diesel engine). The characteristic diesel “knock” occurs when fuel that has been injected into the cylinder ignites after a delay causing a late shock wave. Minimizing this delay results in less unburned fuel in the cylinder and less intense knock. Therefore, higher-cetane fuel usually causes an engine to run more smoothly and quietly. This does not necessarily translate into greater efficiency, although it may in certain engines. The cetane number is determined in a single cylinder. Cooperative Fuel Research (CFR) engine by comparing its ignition quality with that of reference blends of known cetane number.

Cetane number = 0.72 diesel index (10)

Calculated cetane index (CCI) is determined by four variables:

$$\begin{aligned} CCI = 45.2 &+ (0.0892)(T_{10}N) \\ &+ [0.131 + 0.901(B)][T_{50}N] \\ &+ [0.0523 - (0.420)B][T_{90}N] \\ &+ [0.00049][(T_{10}N)^2 - (T_{90}N)^2] \\ &+ 107B + 60B^2 \end{aligned}$$

where T_{10} = 10 % distillation temperature, °C
T_{50} = 50 % distillation temperature, °C
T_{90} = 90% distillation temperature, °C
$B = e^{-3.5DN} - 1$
D = Density @ 15°C
DN = D – 0.85

CFR: Combined feed ratio. The ratio of total feed (including recycle) to fresh feed.

CGO: Coker gas oil.

Charge Capacity: The input (feed) capacity of the refinery processing facilities.

Characterization Factor (CF): **1**. An index of feed quality, also useful for correlating data on physical properties. The Watson or Universal Oil Property (UOP) characterization factor K_W is defined as the cube root of the mean average boiling point in °R divided by the specific gravity. An indication of carbon-to-hydrogen ratio. K_w is expressed by

$$K_W = \frac{T_B^{1/3}}{Sp.Gr}$$

where
T_B = mean average boiling point, °R [°F + 460]
Sp.Gr = Specific gravity at 60°F

T_B is the average boiling point in °R taken from five temperatures corresponding to 10, 30, 50, 70, and 90% volume vaporized.

2. A calculated factor used to correlate properties for petroleum streams. It is a measure of the paraffinicity of the stream and is defined as $CF = \left[MABP^{1/3}/Sp.Gr\right]$, where MABP = mean average boiling point temperature, °R (=460 + °F), and Sp.Gr. = specific gravity at 60°F (15.9°C) relative to water.

Typically, Watson characterization factor varies between 10.5 and 13 for various crude streams. Highly paraffinic crude typically possesses a K_w of 13. On the other hand, highly naphthenic crude possesses a K_w factor of 10.5. Therefore, the Watson characterization factor can be used to judge the quality of the crude oil in terms of the dominance of the paraffinic or naphthenic compounds.

Checklist: A detailed list of desired systems attributes for a facility. It is used to assess the acceptability of a facility compared to accepted norms.

Clarified Oil: The heaviest stream from a catalytic cracking process after settling to remove suspended catalyst particles.

Clear Treating: An elevated temperature and pressure process usually applied to thermally cracked naphthas to improve stability and color. The stability is increased by the adsorption and polymerization of reactive diolefins in the cracked naphtha. Clay treating is used for treating jet fuel to remove surface agents that adversely affect the water separator index specifications.

Clear: Without lead. Federal regulations require that fuels containing lead must be dyed.

Cloud Point: **1**. The temperature at which solidifiable compounds (wax) present in the sample begin to crystallize or separate from the solution under a method of prescribed chilling. **2**. The temperature at which a noticeable cloud of crystals or other solid materials appears when a sample is cooled under prescribed conditions. Cloud point is a typical specification of middle distillate fuels; ASTM D-2500.

Cold Filter Plugging: Is defined as the temperature at which a fuel suspension fails to flow through a standard filter when cooled as prescribed by the test method.

Coke Drum: A large upright drum used as a receptacle for coke formed in the delayed coking process.

Coke: 1. A product of the coking process in the form of mostly solid, densely packed carbon atoms. **2**. Deposits of carbon that settle on catalysts in cat. crackers, cat. reformers, hydrocrackers, and hydrotreaters, and degrade their effectiveness. 3. A carbonaceous deposit is formed by the thermal decomposition of petroleum.

Coker: A refinery process in which heavy feed such as flasher bottoms, cycle oil from a catalytic cracker, or thermal cracked gas oil is cooked at high temperatures. Cracking creates light oils; coke forms in the reactors and needs to be removed after they fill up.

Coking: A refining process in which petroleum oil is heated destructively such that the heaviest materials are converted to coke. There are two processes: delayed coking and fluid coking, with delayed coking being the most widely used.

Coil: A series of pipes in a furnace through which oil flows and is heated.

Color: It is an indication of the thoroughness of the refining process. This is determined by the Saybolt Chromometer or by Lovibond Tintometer. The Saybolt color of petroleum products test is used for quality control and product identification purposes on refined products having an ASTM color of 0.5 or less.

The ASTM color of petroleum products applies to products having an ASTM color of 0.5 or darker, including lubricating oils, heating oils, and diesel fuel oils.

Pale = 4.5 ASTM color or lighter

Red = Darker than 4.5 ASTM

Dark = Darker than 8.0 ASTM

Compressed Natural Gas: **1**. Natural gas that has been compressed under high pressures (typically between 3000 and 3600 psi and held in a container; expands when released for use as a fuel. **2**. Natural gas is compressed to a volume and density that is practical as a portable fuel supply (even when compressed, natural gas is not a liquid). **3**. Natural gas in its gaseous state that has been compressed. **4**. Natural gas that is under pressure. The pressure reduces the volume occupied for the gas so it can be contained in a smaller vessel.

Compressibility: The volume change of a material when pressure is applied.

Compressibility Factor (Z): **1**. The fractional reduction in the volume of a substance with applied pressure. The compressibility factor is a measure of the compressibility of a gas, Z, and is used as a multiplier to adapt the ideal gas law for non-ideal gases. **2**. The ratio of the actual volume of a gas divided by the volume that would be predicted by the ideal gas law, usually referred to as the "Z" factor.

$$Z = \frac{pV}{RT}$$

where p is the pressure, V is the volume, R is the universal gas constant, and T is the absolute temperature.

Compressible Fluid: A fluid in which the density changes with applied pressure. The compressibility of liquids is negligible in comparison with gases and vapors. The isothermal compressibility of a gas is the change in volume per unit volume or density for a unit change in applied pressure given as:

$$c = \frac{-1}{V}\left(\frac{\partial V}{\partial p}\right)_T = \frac{-1}{\rho}\left(\frac{\partial \rho}{\partial p}\right)_T$$

Isothermal compressibility coefficients are frequently used in oil and gas engineering, transient fluid flow calculation, and the determination of the physical properties of substances.

Compression Ratio: Is a measure of the amount of compression that takes place in an engine's cylinder. The ratio of volumes in an internal combustion cylinder when the piston is at the bottom of the stroke to that when the piston is at the top of the stroke, giving a measure of how much the air or air/fuel mixture is compressed in the compression stroke.

$$CP = \frac{V_1}{V_2} = \frac{\text{Volume when piston is @ bottom of stroke}}{\text{Volume when piston is @ top of stroke}}$$

Compressor: **1**. A device that increases the pressure of the gas. Commonly used as a production rate increaser by increasing the gas pressure delivered from low-pressure gas wells to enter the pipeline. The intake into the compressor lowers the wellhead pressure, creating a larger drawdown. **2**. An engine used to increase the pressure of natural gas so that it will flow more easily through a pipeline. **3**. A thermodynamic machine that increases the pressure of a gas flow using mechanical energy. **4**. A mechanical device used to raise the pressure of a gas. Compressors can be of three types: ***axial, centrifugal, or reciprocating***. The usual means of providing the required power are electrical motors, steam turbines, or gas turbines.

Compressor Station: **1**. A booster station is associated with a gas pipeline that uses compressors to increase the gas pressure. When gas turbines are used to provide compressor power, stations can use some of the gas moving through the line as fuel. **2**. Stations located along natural gas pipelines that recompress gas to ensure an even flow.

Condensation: Reaction in which aromatic ring structures combine to form ring structures larger than the reactants.

Condensate: **1**. The relatively small amount of liquid hydrocarbon, typically C_4, s, through naphtha or gas oil that gets produced in the oil patch with unassociated gas. **2**. The liquid formed when a vapor cools.

Conradson Carbon: A test used to determine the amount of carbon residue left after the evaporation and pyrolysis of oil under specified conditions. Expressed as weight percentage; ASTM D-189.

Conradson Carbon Residue (CCR): Results from ASTM test D189. It measures the coke-forming tendencies of oil. It is determined by destructive distillation of a sample to elemental carbon (coke residue), in the absence of air, expressed as the weight percentage of the original sample. A related measure of the carbon residue is called ***Ramsbottom carbon residue***. A crude oil with a high CCR has a low value as refinery feedstock.

Conradson Carbon Residue (ASTM D 1289): ASTM D 4530 microcarbon residue: This procedure determines the carbon residue left after the evaporation and pyrolysis of an oil sample under prescribed conditions and is a rough indicator of oil's relative coke-forming tendency or the contamination of a lighter distillate fraction with a heavier distillate fraction or residue.

Carbon residue and atomic H-to-C ratio are correlated by:

$$H/C = 171 - 0.015CR \text{ (Conradson)}$$

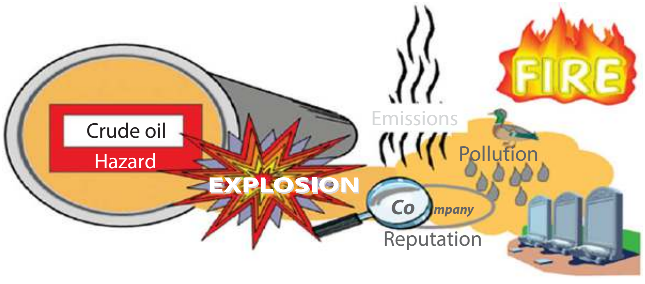
Consequence: **1**. Is the ultimate harm that may occur due to a credible hazard release scenario. **2**. The direct undesirable result of an accident sequence usually involving a fire, explosion, or the release of toxic material. Consequence description may include estimates of the effects of an accident in terms of factors such as health impacts, physical destruction, environmental damage, business interruption, and the public reaction of company prestige (see Figure 6).

Continuous Catalytic Reforming (CCR) process: Continuous catalytic reforming process occurs where the catalyst is circulated through the reactors and a regeneration step, analogous to catalytic cracking processes.

Continuous Stirred Tank Reactor (CSTR): **1**. A type of idealized chemical reactor that is used to contain a chemical reaction in which liquid reactants continuously flow into the reactor and products are continuously removed such that there is no accumulation within the reactor. By assuming perfect mixing of the reactants within the reactor, by using a stirrer/mixer, the composition of the material is therefore assumed to be the same as the composition at all points within the reactor. **2**. Reactors are characterized by a continuous flow of reactants into and a continuous flow of products from the reaction system. Examples are the plug flow reactor and the continuous stirred flow reactor (Figure 7).

Control of Major Accident Hazards (COMAH): The legislation requires that businesses holding more than threshold quantities of named dangerous substances "Take all necessary measures to prevent major accidents involving dangerous substances. Limit the consequences to people and the environment of any major accidents which do occur." Plant designers need to consider whether their proposed plant will be covered by this legislation at the earliest stages.

Control of Substances Hazardous to Health (COSHH): The legislation that requires risk assessment and the control of hazards associated with all chemicals and used in a business that has potentially hazardous properties. A consideration of the properties of chemicals used as feedstock, intermediates, and products is a basic part of plant design. Inherently safe design requires us to consider these issues at the earliest stage.



Resulting event or chain of events

Figure 6 A consequence.

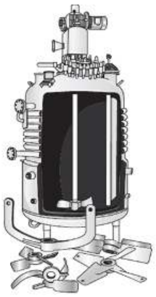


Figure 7 (a) Continuous stirred tank reactor with different impeller types. (b) A series of continuous-flow stirred tank reactors.

Conversion: **1**. A measure of the completeness of a chemical reaction. It is often presented as the fraction of a particular reactant consumed by the chemical reaction. The ***conversion per pass*** is a measure of the limiting reactant that is converted in a chemical reactor and recycled in combination with fresh reactant feed. Not all reactions are complete within the reactor, and in many cases, unreacted reactants are separated from products and recycled for further action. **2**. Typically, the fraction of a feedstock is converted to gasoline and lighter components.

Correlation Index (CI): The US Bureau of Mines factor for evaluating individual fractions from crude oil. The CI scale is based upon straight chain hydrocarbons having a CI value of 0 and benzene having a value of 100. The lower the CI value, the greater the concentrations of paraffin hydrocarbons in the fraction, and the higher the CI value, the greater the concentrations of naphthenes and aromatics. CI is an indication of the hydrocarbon-to-carbon ratio and the aromaticity of the sample. CI is expressed by:

$$\mathrm{CI} = \frac{87552}{\mathrm{T_B}} + 473.7\,\mathrm{Sp.Gr} - 456.8$$

$$\mathrm{CI} = \frac{48640}{\mathrm{K}} + 473.7\mathrm{d} - 456.8$$

where

d = specific gravity at 15°C/15°C

K = average boiling point, (K = °C + 273.15)

T_B = mean average boiling point, °R

Sp.Gr = specific gravity at 60°F

Corrosion: **1**. The deteriorating chemical reaction of a metal with the fluids with which it is in contact. **2**. The gradual decomposition or destruction of a material by chemical action, often due to an electrochemical reaction. Corrosion may be caused by (a) stray current electrolysis, (b) galvanic corrosion caused by dissimilar metals, and (c) differential-concentration cells. Corrosion starts at the surface of a material and proceeds inward.

Corrosion Inhibition: Corrosion can be defined as the unwanted production of a salt from a metal. Adding acid or oxygen is a good way for this to occur. The main way of slowing corrosion down (inhibition) is by providing an impermeable coating to stop the chemical reaction from occurring in the first place or by providing a more easily attacked metal that will be consumed first (a “sacrificial anode”).

Corrosion Inhibitor: **1**. A chemical substance or combination of substances that, when present in the environment, prevents or reduces corrosion. **2**. A substance that slows the rate of corrosion.

Corrosive Gas: **1**. A gas that attacks metal or other specified targets. Most commonly CO_2 and H_2S. Usually in association with water or water vapor. Oxygen can be described as a corrosive gas in some cases. **2**. In water, dissolved oxygen reacts readily with metals at the anode of a corrosion cell, accelerating the rate of corrosion until a film of oxidation products such as rust forms. At the cathode where hydrogen gas may form a coating on it and therefore slows the corrosion rate, oxygen reacts rapidly with hydrogen gas, forming water, and again increases the rate of corrosion.

Cracking: The breaking down of higher-molecular-weight hydrocarbons to lighter components by the application of heat. Cracking in the presence of a suitable catalyst produces an improvement in yield and quality over simple thermal cracking.

Cracking Correction: Correction to a laboratory distillation to account for the lowering of the recorded temperatures because of the thermal cracking of the sample in the distillation flask. Cracking occurs for most petroleum stocks at temperatures greater than about 650°F (344°C) at atmospheric pressure.

Cracked Stock: A petroleum stock that has been produced in a cracking operation, either catalytic or thermal. Cracked stocks contain hydrogen-deficient compounds such as olefins (C_nH_{2n}) and aromatics (C_nH_{2n-6}).

Critical Point: The temperature and pressure at which a component or mixture of components enters a dense phase, being neither liquid nor vapor.

Critical Pressure: The vapor pressure at the critical temperature.

Critical Temperature: The temperature above which a component cannot be liquefied. For mixtures, the temperature above which all of the mixture cannot be liquefied.

Crude Assay Distillation: See Fifteen-Five (15/5) Distillation.

Crude Chemistry: Fundamentally, crude oil consists of 84–87 wt% carbon, 11–14% hydrogen, 0–3 wt% sulfur, 0–2 wt% oxygen, 0–0.6 wt% nitrogen, and metals ranging from 0 to 100 ppm. Understanding thoroughly the fundamentals of crude chemistry is very important in various refining processes. The existence of compounds with various functional groups and their dominance or reduction in various refinery products are what are essentially targeted in various chemical and physical processes in the refinery.

Based on chemical analysis and the existence of various functional groups, refinery crude can be broadly categorized into about nine categories:

1. Paraffins, C_nH_{2n+2}, CH_4, C_2H_6, C_3H_8	**4**. Aromatics, C_nH_{2n-6}, C_6H_6, C_7H_8, C_8H_{10}	**7**. Oxygen-containing compounds, R-OH, CH_3OH, C_6H_5OH
2. Olefins, C_nH_{2n}. C_2H_4, C_3H_6	**5**. Naphthalene	**8**. Resins
3. Naphthenes, C_nH_{2n}, C_6H_{12},	**6**. Organic sulfur compounds, RSH, CH_3SH, R – S – R′	**9**. Asphaltenes

Crude and Crude Oil: **1**. A range of principally carbon–hydrogen chain compounds with generally straight carbon chain lengths of C_1 (methane) to C_{60+}, compounds boiling higher than 2000°F (1094°C). The straight-chain materials are alkanes. **2**. Oil as it comes from the well; unrefined petroleum. **3**. The petroleum liquids as they come from the ground; formed from animal and vegetable material that is collected at the bottom of ancient seas. **4**. A tarry group consisting of mixed carbon compounds with a highly variable composition. **5**. A mixture of hydrocarbons that exists in the liquid phase in natural underground reservoirs and remains liquid at atmospheric pressure after passing through surface-separating facilities. Depending upon the characteristics of the crude stream, it may also include the following:

- Small amounts of hydrocarbons that exist in the gaseous phase in natural underground reservoirs but are liquid at atmospheric pressure after being removed from oil well (casing head) gas in lease separators and are subsequently commingled with the crude stream without being separately measured. Lease condensate recovered as a liquid from natural gas wells in lease or field separation facilities and later mixed into the crude stream is also included.
- Small amounts of non-hydrocarbons produced from oil, such as sulfur and various metals.
- Drip gases and liquid hydrocarbons produced from tar sands, gilsonite, and oil shale.
- Liquid produced at natural gas processing plants is excluded. Crude oil is refined to produce a wide range of petroleum products, including heating oils; gasoline, diesel, and jet fuels; lubricants; asphalt; ethane, propane, and butane; and many other products used for their energy or chemical content.

The basic types of crudes are asphalt, naphthenic, or paraffinic depending on the relative proportion of these types of hydrocarbons present.

Crude Oil Assay: Is a precise and detailed analysis of carefully selected samples of crude thoroughly representative of average production quality. It helps to assess the potential sales value of new crude oil and to plan for its most effective utilization. Numerous important feed and product characterization properties in refinery engineering include:

1. API gravity
2. Watson characterization factor
3. Viscosity
4. Sulfur content
5. True boiling point (TBP) curve
6. Pour point
7. Flash and fire point
8. ASTM distillation curve
9. Octane number

Crude Oil Losses: This represents the volume of crude oil reported by petroleum refineries as being lost in their operations. These losses are due to spills, contamination, fires, etc., as opposed to refinery processing losses.

Crude Oil Production: The volume of crude oil produced from oil reservoirs during given periods of time. The amount of such production for a given period is measured as volumes delivered from lease storage tanks, (i.e., the point of custody transfer) to pipelines, trucks, or other media for transport to refineries or terminals with adjustments for (1) net differences between opening and closing lease inventories and (2) basic sediment and water (BS & W).

Crude Oil Qualities: This refers to two properties of crude oil, the sulfur content and API gravity, which affect processing complexity and product characteristics.

Cryogenics: The production and application of low-temperature phenomena. The cryogenic temperature range is usually from -238°F (-150°C) to absolute zero -460°F (-273°C), the temperature at which molecular motion essentially stops. The most important commercial application of the cryogenic gas liquefaction technique is the storage, transportation, and regasification of LNG.

Cryogenic Liquid or Cryogenics: Cryogenic liquids are liquefied gases that are kept in their liquid state at very low temperatures and have a normal boiling point below -238°F (-150°C). All cryogenic liquids are gases at normal temperatures and pressures. These liquids include methane (CH_4), oxygen (O_2), nitrogen (N_2), helium (He), and hydrogen (H_2). Cryogens are normally stored at low pressures.

Cryogenic Recovery: Cryogenic recovery processes are carried out at temperatures lower than -150°F (-101°C). The low temperatures allow the plant to recover over 90% of the ethane in the natural gas. Most new gas processing plants use cryogenic recovery technology.

CSB: An acronym for the US Chemical Safety and Hazard Investigation Board. An agency of the US government chartered to investigate chemical industry incidents, determine their root causes, and publish their findings to prevent similar incidents from occurring.

Cut: A portion of crude oil boiling within certain temperature limits. Usually, the limits are on a crude assay true boiling point (TBP) basis.

Cut Point Temperature, Cut Points: A temperature limit of a cut, usually on a true boiling point (TBP) basis, although ASTM distillation cut point is not uncommon. The boiling point curve most commonly used to define cut points is the TBP at one atmosphere of pressure.

Cut Point Ranges: A series of cut point temperatures are defined for a petroleum stock. The cut point ranges are the temperature differences between adjacent cut point temperatures. When developing petroleum pseudo-components for a petroleum stock, cut point ranges must be defined that include the total boiling point range of the stock.

Cutter Stock: Diluent added to residue to meet residual fuel specifications for viscosity and perhaps sulfur content. Typically cracked gas oil.

Cycloparaffin: A paraffin molecule with a ring structure.

Cycle Oil, Cycle Stock: An oil stock, containing a hydrogen-deficient compound that was produced in a thermal or catalytic cracking operation.

Cyclization: Chemical reaction in which a non-ring structure paraffin or olefins are converted into ring structures.

Cyclo-olefins: Unsaturated ring structure with one or two double bonds in the ring.

Darcy–Weisbach Equation: An equation used in fluid mechanics to determine the pressure or head loss due to friction within a straight length of pipe for a flowing fluid. The frictional pressure drop, Δp_f (psi, bar) is expressed by

$$\Delta p_f = f_D \left(\frac{L}{d}\right)\frac{\rho v^2}{2}$$

where

$$f_D = \left(\frac{\tau_w}{\frac{\rho v^2}{2}}\right)$$

In the form of frictional head loss, h_f, (ft, m) is:

$$h_f = f_D \left(\frac{L}{d}\right)\frac{v^2}{2g}$$

where τ_w is the shear stress, f_D is the Darcy friction factor, dimensionless ($f_D = 4f_F$), f_F is the fanning friction factor, L and d are the pipe length (ft, m) and inside diameter (ft., m), v is the average velocity of the fluid (ft), ρ is the fluid density (lb_m/ft^3, kg/m^3), and g is the acceleration due to gravity (ft/s^2, m/s^2). It is known as the Darcy–Weisbach or Moody friction fawctor, whose value depends on the nature of the flow and surface roughness of the pipe. This Darcy friction factor is four times the Fanning friction factor. The value of the friction factor can be determined from various empirical equations and published charts such as the Moody diagram (See Figure 8).

An empirical equation known as the Colebrook–White equation has been proposed for calculating the friction factor in the turbulent flow:

$$\frac{1}{\sqrt{f_D}} = -2\log_{10}\left(\frac{e}{3.7D} + \frac{2.51}{Re\sqrt{f_D}}\right)$$

where
D = pipe inside diameter, in
e = absolute pipe roughness, in
Re = Reynolds number, dimensionless

The term $f_D(L/d)$ may be substituted with a head loss coefficient K (also known as the resistance coefficient) and then becomes

$$h_f = K\frac{v^2}{2g}$$

The head loss in a straight piece of pipe is represented as a multiple of the velocity head $v^2/2g$. Following a similar analysis, we can state that the pressure drop through a valve or fitting can be represented by $K(v^2/2g)$, where the coefficient K is specific to the valve and fitting. Note that this method is only applicable to turbulent flow through pipe fittings and valves. Recently, K is presented by Hooper's 2-K method and Darby's 3-K method.

DAO - Deasphalted oil: The raffinate product from the propane deasphalting unit.

D1160: ASTM laboratory distillation method for high-boiling streams. The D1160 is performed under vacuum conditions with 10 mm Hg being the most common pressure used for the test. D1160 data are normally reported at a 760 mm Hg basis.

D2887: ASTM-simulated distillation method for high-boiling streams. The D2887 has an upper limit of 1000°F (538°C), and the temperatures are reported versus weight percent distilled. A normal paraffin standard is used to convert the chromatographic results to a boiling point curve.

D3710: ASTM-simulated distillation method for gasoline and light naphthas. D3710 data are reported on a volume basis.

D86: ASTM laboratory distillation method conducted at atmospheric pressure for streams boiling below approximately 700°F (371°C). The D86 is the most commonly used laboratory distillation for petroleum stocks.

Deasphalting: Process for removing asphalt from petroleum fractions, such as reduced crude.

Debottlenecking: **1**. Increasing the production capacity of existing facilities through the modification of existing equipment to remove throughput restrictions. Debottlenecking generally increases capacity for a fraction of the cost of building new facilities. **2**. The process of increasing the production capacity of existing facilities through the modification of existing equipment to remove throughput restrictions. **3**. A program, typically in surface facilities and lines, to remove pressure drop causing flow restrictions.

Debutanizer: A column that removes n-butanes (nC_4H_{10}) and lighter in the top product.

Decant Oil: The bottom stream from the FCC unit distillation tower after the catalyst has been separated from it.

Decanted Water: Insoluble water that is drawn from a drum containing condensed hydrocarbons and water.

Decarbonization: A process of reducing the amount of carbon dioxide released into the atmosphere.

Decoking: The process of removing coke from catalysts in a catalytic cracker, catalytic reformer, hydrocracker, or hydrotreaters. Usually, heated air will oxidize the coke to carbon monoxide or carbon dioxide.

Deethanizer: A column that removes ethane (C_2H_6) and lighter in the top product.

Deflagration (i.e., "to burn down"): Is a term describing subsonic combustion propagation through heat transfer; hot burning material heats the next layer of cold material and ignites it. Most "fire" found in daily life, from flames to explosions, is deflagration. Deflagration is different from detonation, which propagates supersonically through shock waves.

Delayed Coker: A process unit in which residue is cooked until it cracks to coke and light products.

Delayed Coking: **1**. A semi-continuous thermal process for the conversion of heavy stock to lighter material. The method involves pre-heating the feedstock in a pipe still, discharging it into large insulated coke drums, and retaining it there for a particular length of time for cracking to occur. Gas, gasoline, and gas oil are recovered as overhead products, and finally, coke is removed. **2**. A process by which heavier crude oil fractions can be thermally decomposed under conditions of elevated temperatures and low pressure to produce a mixture of lighter oils and petroleum coke. The light oils can be processed further in other refinery units to meet product specifications. The coke can be used either as a fuel or in other applications such as the manufacturing of steel or aluminum.

Dehydrogenation: A chemical reaction in which a compound loses bonded hydrogen.

Deisobutanizer: A column that removes isobutane (iC_4H_{10}) and lighter in the top product.

Demethanizer: A column that removes methane (CH_4) and lighter in the top product.

Density: The density of crude oil and petroleum fractions is usually specified in °API, specific gravity, or kilograms per cubic meter (kg/m^3). The numerical values of specific gravity and kg/m^3 are equal; that is, a fraction with a specific gravity of 0.873 has a density of 873 kg/m^3. The API scale runs opposite to that of specific gravity, with larger values for less dense materials and smaller values for more dense fractions (water = 10°API). By definition, °API is always 60°F (15.6°C) for a liquid.

Depentanizer: A column that removes n-pentane (nC_5H_{12}) and lighter in the top product.

Depropanizer: A column that removes propane (C_3H_8) and lighter in the top product.

Desalting: A process that removes chlorides and other inorganic salts from crude oil by injecting water and applying an electrostatic field to force the salt into the aqueous phase.

Desiccant: An absorbent or adsorbent, liquid or solid, that removes water or water vapor from an air stream.

Desiccant Drying: The use of a drying agent to remove moisture from a stream of oil or gas. In certain product pipelines, great effort is made to remove all the water vapor before putting the line into service. To accomplish this, desiccant dried air or inert gas is pumped through the line to absorb the moisture that may be present even in the ambient air in the line.

Desiccation: The process of drying and removing the moisture within a material. It involves the use of a drying agent known as a desiccant. Desiccants that function by the adsorption of moisture include silica gel and activated alumina, while chemical desiccants that function by the reaction with water to form hydrates include calcium chloride and solid sodium hydroxide. A desiccator is a container used for drying substances or for keeping them dry and free of moisture. Laboratory desiccators are made of glass and contain a drying agent such as silica gel.

Design Codes (design standards): Published standards required for equipment and working practices within the chemical and process industries that represent good practice and define the level of standard of design. Developed and evolved over many years and based on tried and tested practices. There are a number of national standards organizations and institutions that provide published standards for design, materials, fabrication, the testing of processes, and equipment. These include the American Petroleum Institute (API), the American National Standards Institute (ANSI), the American Society of Mechanical Engineers (ASME), the American Society for Testing and Materials (ASTM), the American Iron and Steel Institute (AISI), and the British Standards Institute (BSI).

Desorption: The release of materials that have been absorbed or adsorbed in or onto a formation.

Desulfurization: The removal of sulfur, from molten metals, petroleum oil, or flue gases. Petroleum desulfurization is a process that removes sulfur and its compounds from various streams during the refining process. Desulfurization processes include catalytic hydrotreating and other chemical/physical processes such as adsorption. Desulfurization processes vary based on the type of stream treated (e.g., naphtha, distillate, heavy gas oil) and the amount of sulfur removed (e.g., sulfur reduction to 10 ppm). See Catalytic Hydrotreating.

Desuperheating zone: A section of a distillation/fractionating column where a superheated vapor is cooled and some liquid is condensed. FCC main fractionators have a desuperheating zone.

Detonation ("to thunder down"): Is a type of combustion involving a supersonic exothermic front accelerating through a medium that eventually drives a shock front propagating directly in front of it. Detonations occur in both conventional solid and liquid explosives, as well as in reactive gases. The velocity of detonation in solid and liquid explosives is much higher than that in gaseous ones, which allows the wave system to be observed with greater detail.

An extraordinary variety of fuels may occur as gases, droplet fogs, or dust suspensions. Oxidants include halogens, ozone, hydrogen peroxide, and oxides of nitrogen. Gaseous detonations are often associated with a mixture of fuel and oxidant in a composition somewhat below the conventional flammability ratio. They happen most often in confined systems, but they sometimes occur in large vapor clouds. Other materials, such as acetylene, ozone, and hydrogen peroxide are detonable in the absence of oxygen. See Knocking.

Dewaxing: The removal of wax from lubricating oils, either by chilling and filtering solvent extraction or selective hydrocracking.

Dew Point: **1**. A vapor at its dew point temperature is on the verge of starting to condense to a liquid. Cool the vapor by 1°F, or raise its pressure by 1 psi and it will form drops of liquid. Air at 100% relative humidity is at its dew point temperature. Cool it by 1°F and it starts to rain. **2**. The temperature and pressure at which the first drop of liquid will condense for a component or a mixture of components. **3**. The temperature at a given pressure at which vapor will form the first drop of liquid on the subtraction of heat. Further cooling of the liquid at its dew point results in the condensation of a part or all of the vapors as a liquid. **4**. The temperature at which vaporized materials start to condense into liquid form. **5**. The temperature at which liquids begin to condense from the vapor phase in a gas stream.

Diene: Same as diolefin.

Diesel: **1**. An internal combustion engine in which ignition occurs by injecting fuel in a cylinder where the air has been compressed and is at a very high temperature, causing self-ignition. **2**. Distillate fuel used in a diesel engine. (*See the Diesel engine*).

Diesel Fuel: A fuel produced for diesel engines with a typical ASTM 86 boiling point range of 450–675°F (233–358°C).

Diesel Index (DI): A measure of the ignition quality of diesel fuel. Diesel index is defined as

$$\mathrm{DI} = \frac{(^\circ\mathrm{API})(\mathrm{Aniline\ Point})}{100}$$

The higher the diesel index, the more satisfactory the ignition quality of the fuel. By means of correlations unique to each crude and manufacturing process, this quality can be used to predict the cetane number (if no standardized test for the latter is available).

Diolefin: C_nH_{2n}: Paraffin-type molecule except that it is missing hydrogen atoms, causing it to have two double bonds somewhere along the chains.

DIPE: Di-isopropyl ether. An oxygenate is used in motor fuels.

Disposition: The components of petroleum disposition are stock change, crude oil losses, refinery inputs, exports, and products supplied for domestic consumption.

Distillate Fuel Oil: A general classification for one of the petroleum fractions produced in conventional distillation operations. It includes diesel fuels and fuel oils. Products known as No. 1, No. 2, and No. 4 diesel fuels are used on highway diesel engines, such as those in trucks and automobiles, as well as off-highway engines, such as those in railroad locomotives and agricultural machinery. Products known as No.1, No. 2, and No. 4 fuel oils are used primarily for space heating and electric power generation.

No. 1 Distillate. A light petroleum distillate that can be used as either diesel fuel or fuel oil.

No. 1 Diesel Fuel. A light distillate fuel oil that has distillation temperatures of 550°F (288°C) at the 90% point and meets the specifications defined in ASTM Specification D 975. It is used in high-speed diesel engines generally operated under frequent speed and load changes, such as those in city buses and similar vehicles.

No. 1 Fuel Oil. A light distillate fuel oil that has distillation temperatures of 400°F (204°C) at the 10% recovery point and 550°F (288°C) at the 90% point and meets the specifications defined in ASTM Specifications D 396. It is used primarily as a fuel for portable outdoor stoves and portable outdoor heaters.

No. 2 Distillate: A petroleum distillate that can be used as either diesel fuel or fuel oil.

No. 2 Diesel Fuel: A fuel that has a distillation temperature of 500°F (260°C) at the 10% recovery point and 640°F (338°C) at the 90% recovery point and meets the specifications defined in ASTM Specification D 975. It is used in high-speed diesel engines that are generally operated under uniform speed and load conditions, such as those in railroad locomotives, trucks, and automobiles.

Low Sulfur No. **2 Diesel Fuel**. No. 2 diesel fuel that has a sulfur level no higher than 0.05% by weight. It is used primarily in motor vehicle diesel engines for on-highway use.

High Sulfur No. 2. **Diesel Fuel**. No. 2 diesel fuel that has a sulfur level above 0.05% by weight.

No. 2 Fuel Oil (Heating Oil): A distillate fuel oil that has distillation temperatures of 400°F (204°C) at the 10% recovery point and 640°F (338°C) at the 90% recovery point and meets the specifications defined in ASTM Specification D 396. It is used in atomizing-type burners for domestic heating or for moderate-capacity commercial/industrial burner units.

No. 4 Fuel. A distillate fuel oil made by blending distillate fuel oil and residual fuel oil stocks. It conforms with ASTM Specification D 396 or Federal Specification VV–F-815C and is used extensively in industrial plants and in commercial burner installations that are not equipped with preheating facilities. It also includes No. 4 diesel fuel used for low- and medium-speed diesel engines and conforms to ASTM Specification D975.

No. 4 Diesel Fuel. See No. 4 Fuel

No. 4 Fuel Oil. See No. 4 Fuel.

Distillate: **1**. The liquid obtained by condensing the vapor given off by a boiling liquid. **2**. Any stream except the bottoms coming from a fractionator. **3**. The products or streams in the light gas oil range such as straight-run light gas oil, cat. cracked light gas oil, heating oil, or diesel.

Distillation: Same as fractionation. A separation process that results in separated products with different boiling ranges. Distillation is carried out in a way that the materials being separated are not subjected to conditions that would cause them to crack or otherwise decompose or chemically change. It is a physical process.

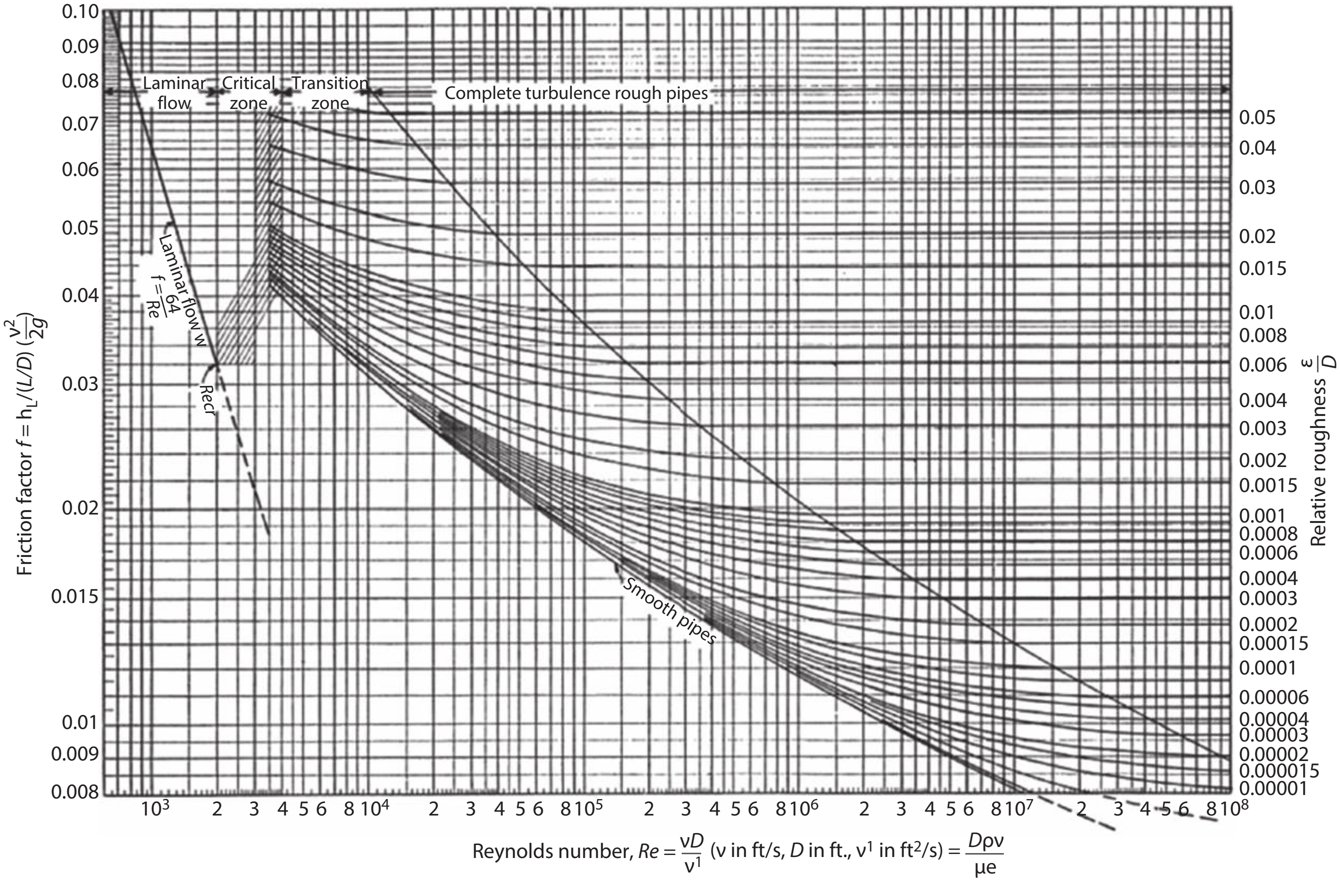


Figure 8 Moody diagram.

Distillation Column: A tall vertical cylindrical vessel used for the process of distillation. Hot vapor rises up the column, which is brought into intimate contact with cooled liquid descending on stages or trays for a sufficient period of time so as to reach equilibrium between the vapor and the liquid. The vapor rises up from the tray below through perforations in the tray, and the liquid on the tray flows over a weir to the tray below. In this way, the more volatile component increases in concentration progressively up the column. In continuous distillation, fresh feed is admitted at the tray corresponding to the same composition. Below the feed point, the section of a column is known as the stripping section, while the above is referred to as the rectifying section. A reboiler heat exchanger is used to boil the bottom product and produce vapor for the column. A condenser is used to condense some or all of the vapor from the top of the column. A small portion of liquid is returned to the column as reflux. The height of the column is an indication of the ease or difficulty of separation. For example, an ethylene splitter in a refinery used to separate ethylene from ethane, which has close boiling points, requires many trays, and the column is very tall. The width of the column is an indication of the internal vapor and liquid rates (See Figure 9).

Distillation Curves: In addition to True Boiling Point (TBP) or good fractionation distillations, there are at least three other major types of distillation curves or ways of relating vapor temperature and percentage vaporized: (a) equilibrium flash vaporization, (b) ASTM or non-fractionating distillations, and (c) Hempel or semi-fractionating distillations (See Figure 10).

Distillation Range: See boiling range.

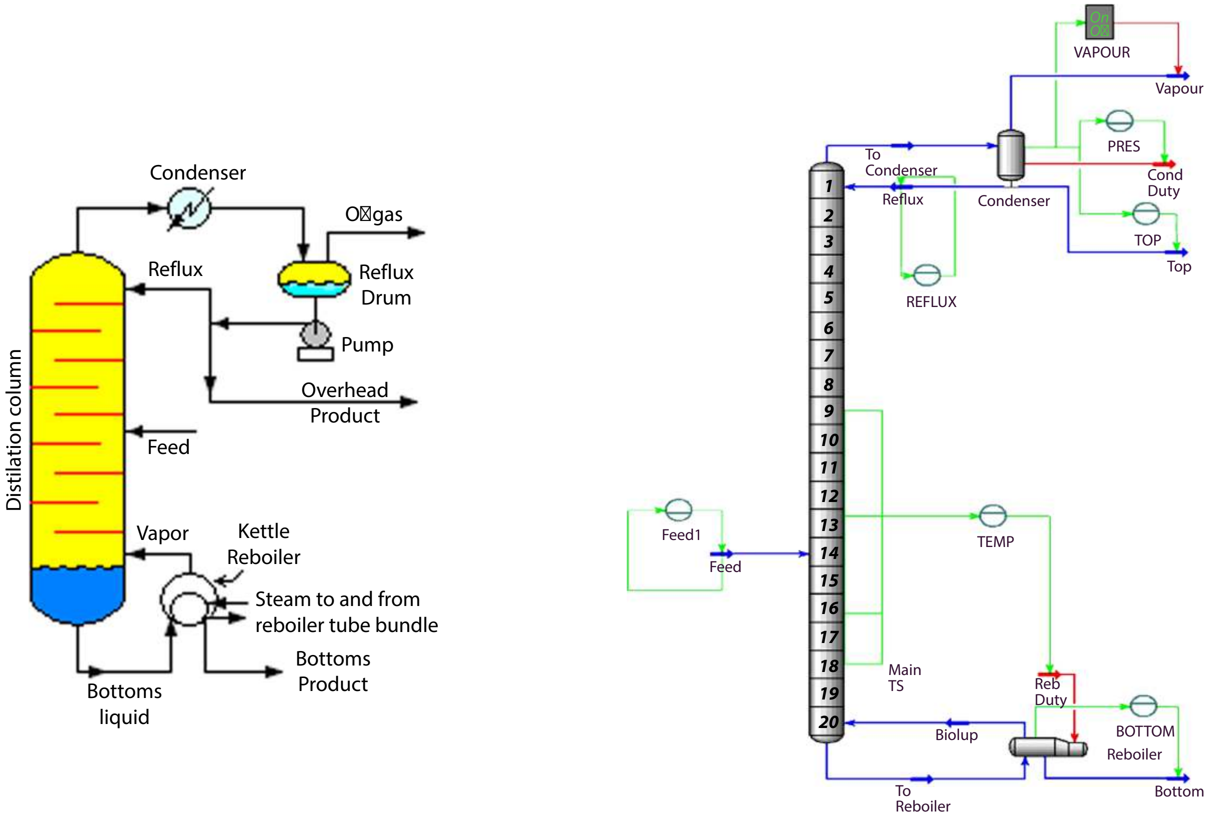


Figure 9 Diagrams of a distillation column.

Distillation Train: A sequence of distillation columns used to separate components from a multicomponent feed. Each column is required to perform a particular separation of either a pure component or a cut between two components. For example, in the separation of four components, ABCD in a mixture in which A is the most volatile and D is the least; then, the five possible separation sequences requiring three columns are:

Separation	Column 1	Column 2	Column 3
1	A:BCD	B:CD	C:D
2	A:BCD	BC:D	B:C
3	AB:CD	A:B	C:D
4	ABC:D	A:BC	B:C
5	ABC:D	AB:C	A:B

Where it is required to separate a larger number of components, the number of possible separation sequences becomes much larger according to the relationship

$$N = \frac{(2n-2)!}{n!(n-1)!}$$

where N is the number of sequences and n is the number of components:

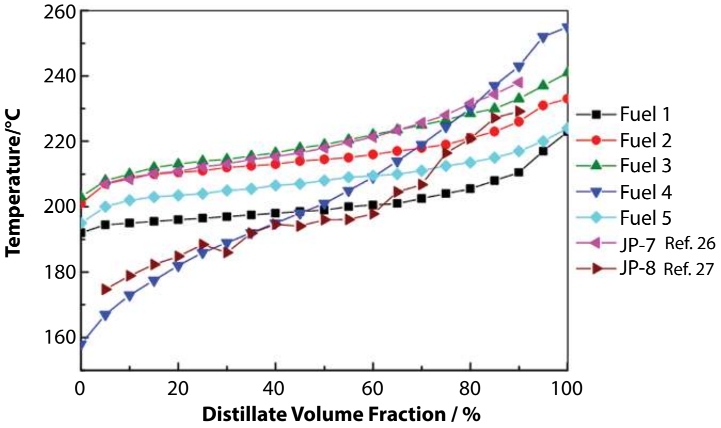


Figure 10 A typical distillation curve of petroleum products.

Components (n)	4	5	6	7	8	9	10
Sequences (N)	5	14	42	132	429	1430	4862

Distributed Component: A component that appears in both the top and bottom products from a distillation/fractionating column separating zone.

Distributed Control System: A system that divides process control functions into specific areas interconnected by communications (normally data highways) to form a single entity. It is characterized by digital controllers and typically by central operation interfaces.

Distributed control systems consist of subsystems that are functionally integrated but may be physically separated and remotely located from one another. Distributed control systems generally have at least one shared function within the system. This may be the controller, the communication link, or the display device. All three of these functions may be shared.

A system of dividing plant or process control into several areas of responsibility, each managed by its own Central Processing Unit, in which the whole are interconnected to form a single entity usually by communication buses of various kinds.

Distributor: A device in a vessel that disperses either liquid or vapor to promote better circulation.

Doctor Test: A method for determining the presence of mercaptan sulfur petroleum products. This test is used for products in which a “sweet” odor is desirable for commercial reasons, especially naphtha; ASTM D-484.

Dow Fire and Explosion Index (F & EI): A method (developed by Dow Chemical Company) for ranking the relative fire and explosion risk associated with a process. Analysts calculate various hazard and explosion indexes using material characteristics and process data.

Downcomer: A device to direct the liquid from a distillation column tray to the next lower distillation tray.

Draw, Side Draw: A product stream withdrawn from a distillation column at a location above the bottom tray and below the top tray. Draws may be vapor or liquid phase.

Dropping Point of Lubricating Greases: Dropping points are used for identification and quality control purposes and can be an identification of the highest temperature of utility for some applications. This is the temperature at which grease passes from a semisolid to a liquid state under prescribed conditions.

Dry Gas: All C_1 to C_3 material, whether associated with crude or produced as a by-product of refinery processing. Convention often includes hydrogen in dry gas yields.

Effective Cut Points: Cut points that can be considered a clean-cut, ignoring any tail ends.

Emergency: A condition of danger that requires immediate action.

Emergency Isolation Valve (EIV): A valve that, in event of a fire, rupture, or loss of containment, is used to stop the release of flammable or combustible liquids, combustible gas, or potentially toxic material. An EIV can be either hand-operated or power-operated (air, hydraulic, or electrical actuation).

Emergency Shutdown (ESD): A method to rapidly cease the operation of a process and isolate it from incoming and outgoing connections or flows to reduce the likelihood of an unwanted event from continuing or occurring. Critical valves shut to isolate sections of the process. Other valves may be opened to depressurize vessels or rapidly discharge the contents of reactors to quench tanks. Emergency shutdowns may occur due to changes in process conditions causing unstable or unsafe operating conditions, a failure in the control system, operator intervention causing unsafe conditions, plant and pipe failure, or some other external event such as an electrical storm or natural catastrophes as earthquakes or coastal flooding.

Emulsion: A colloidal suspension of one liquid dispersed within another. The dispersed phase has droplet sizes usually less than 1 mm. Surfactants or emulsifiers are surface-active agents and are used to stabilize emulsions. In the offshore oil industry, emulsions form at the interface of water and oil in crude oil gravity separators. Sufficient hold-up time is used to separate the emulsion, or, alternatively, surface-active agents are used to encourage separation.

Endothermic reaction: **1**. A chemical reaction that absorbs heat from its surroundings in order for the reaction to proceed. Such reactions have a positive enthalpy change and therefore do not occur spontaneously. **2**. A reaction in which heat must be added to maintain reactants and products at a constant temperature.

E85: Fuel containing a blend of 70–85% ethanol.

End Point (final boiling point): **1**. The highest boiling point recorded for a laboratory distillation. Usually, there is some residual material in the laboratory still, and the end point is not the highest boiling point material in the mixture being distilled. **2**. The lowest temperature at which virtually 100% of petroleum product will boil off to vapor form.

Energy: The capacity or ability of a system to do work. It may be identified by type as being kinetic, potential, internal, and flow or by a source such as electric, chemical, mechanical, nuclear, biological, solar, etc. Energy can be neither created nor destroyed but converted from one form to another. It can be stored as potential energy, nuclear, and chemical energy, whereas kinetic energy is the energy in motion of a body defined as the work that is done in bringing the body to rest. The internal energy is the sum of the potential energy and kinetic energy of the atoms and molecules in the body. Energy as the units Btu, cal, and J.

Energy Balance: **1**. An accounting of the energy inputs and outputs to a process or part of a process, which is separated from the surroundings by an imaginary boundary. All energy forms are included in which the energy input across the boundary must equal the energy output plus any accumulation within the defined boundary. When the conditions are steady and unchanging with time, the energy input is equal to the energy output. The most important energy forms in most processes are kinetic energy, potential energy, enthalpy, heat, and work. Electrical energy is included in electrochemical processes and chemical energy is in processes involving chemical reactions that occur in various reactor types (e.g., batch, continuous stirred tank, plug flow, fixed and catalytic reactors). **2**. The summation of the energy entering a process and the summation of the energy leaving a process. They must be equal for a steady-state process.

Energy Management: Is the planning and operation of energy production and energy consumption units. Objectives are resource conservation, climate protection, and cost savings, while users have permanent access to the energy they need. Energy management is the proactive, organized, and systematic coordination of procurement, conversion, distribution, and the use of energy to meet the requirements, taking into account the environmental and economic objectives. It is also the solution for electric power producers to reduce emissions and improve efficiency and availability. Energy management requires reducing NO_x and greenhouse gas emissions; improving fuel efficiency and reducing SCR operating costs; and streamlining the detection, diagnosis, and remediation of plant reliability, capacity, and efficiency problems. Energy management programs incorporate energy policies, benchmarking, local and corporate goals, the types of energy audits and assessments, reporting systems, and the integration of energy efficiency elements into engineering procedures and purchasing protocols.

Pinch analysis is a tool that is employed in the energy management of chemical facilities and is a methodology for minimizing the energy consumption of chemical processes by calculating thermodynamically feasible energy targets (or minimum energy consumption) and achieving them by optimizing heat recovery systems, energy supply methods, and processing operating conditions. It is also known as process integration, heat integration, energy integration, or pinch technology (See Process Integration).

Engler Distillation: A standard test for determining the volatility characteristics by measuring the percent distilled at various specified temperatures (see ASTM D86).

Engine Knocking (knock, detonation, spark knocking, pinging, or pinking): Spark ignition in internal combustion engines occurs when the combustion of the air/fuel mixture in the cylinder does not start off correctly in response to ignition by the spark plug, but one or more pockets of air/fuel mixture explode outside the envelope of the normal combustion front.

The fuel–air charge is meant to be ignited by the spark plug only and at a precise point in the piston's stroke. Knock occurs when the peak of the combustion process no longer occurs at the optimum moment for the four-stroke cycle. The shock wave creates the characteristic metallic "pinging" sound, and the cylinder pressure increases dramatically. The effects of engine knocking range from inconsequential to completely destructive. See also *Knocking.*

Engineering line diagram (ELD): A diagrammatic representation of a process. Also referred to as ***engineering flow diagram***. It features all process equipment and piping that are required for the start-up and shutdown, emergency, and normal operation of the plant. It also includes insulation requirements; the direction of flows; the identification of the main process and start-up lines; all instrumentation, control, and interlock facilities; the key dimensions and duties of all equipment; the operating and design pressure and temperature for vessels; equipment elevations; set pressures for relief valves; and drainage requirements.

Engineering, Procurement, and Construction Contract: **1**. A legal agreement setting out the terms for all activities required to build a facility to the point that it is ready to undergo preparations for operations as designed. **2**. The final contracting phase in the development of the export portion of the LNG chain that defines the terms under which the detailed design, procurement, construction, and commissioning of the facilities will be conducted. Greenfield LNG project development involves a wide range of design, engineering, fabrication, and construction work far beyond the capabilities of a single contractor. Therefore, an LNG project developer divides the work into a number of segments, each one being the subject of an engineering, procurement, and construction (EPC) contract. **3**. Contract between the owner of a liquefaction plant and an engineering company for the project development and erection. See Front-End Engineering and Design Contract.

Enthalpy (H): The thermal energy of a substance or system with respect to an arbitrary reference point. The enthalpy of a substance is the sum of the internal energy and flow of energy, which is the product of the pressure and specific volume.

$$H = U + pV$$

The reference point for gases is 273 K and for chemical reactions is 298 K.

Enthalpy balance: A form of energy accounting for a process in which the stream energies to and from the process are expressed as enthalpies. At a steady state, the total enthalpy into a process is equal to the total enthalpy out. Where there is an inequality, there is either a loss or an accumulation of material with an associated loss or increase in enthalpy. An enthalpy balance is used to determine the amount of heat that will be generated in the process or that needs to be removed to ensure that the process operates safely and to specification.

Entrainment: A non-equilibrium process by which liquids are mechanically carried into a vapor leaving a process vessel or contacting device.

Entrance and exit losses: The irreversible energy loss caused when fluid enters or leaves an opening, such as into or out of a pipe into a vessel. Where there is a sudden enlargement, such as when a pipe enters a larger pipe or vessel, eddies form and there is a permanent energy loss expressible as a head loss as:

$$H_{exit} = \frac{v^2}{2g}\left(1 - \frac{a}{A}\right)^2$$

where v is the velocity in the smaller pipe, a is the cross-sectional area of the smaller pipe, and A is the cross-sectional area of the larger pipe. For a considerable enlargement, the head loss tends to

$$H_{exit} = \frac{v^2}{2g}$$

With a rapid contraction, it has been found experimentally that the permanent head loss can be given by:

$$H_{exit} = K\frac{v^2}{2g}$$

where for very large contraction, K = 0.5

Entropy (dS): The extent to which the energy in a closed system is unavailable to do useful work. An increase in entropy occurs when the free energy decreases or when the disorder of molecules increases. For a reversible process, entropy remains constant such as in a friction-free adiabatic expansion or compression. The change in entropy is defined as:

$$dS = \frac{dQ}{T}$$

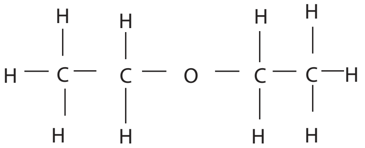
where Q is the heat transferred to or from a system, and T is the absolute temperature. However, all real processes are irreversible, which means that in a closed system, there is a small increase in entropy.

Environmental Protection Agency (EPA), United States: **1**. Governmental agency, established in 1970. Its responsibilities include the regulation of fuel and fuel additives. **2**. The US federal agency that administers federal environmental policies, enforces environmental laws and regulations, performs research, and provides information on environmental subjects. The agency also acts as the chief advisor to the president on American environmental policy and issues. **3**. A federal agency created in 1970 to permit coordinated and effective government action, for the protection of the environment by the systematic abatement and control of pollution, through integration of research monitoring, standard setting, and enforcement activities. **4**. US pollution control enforcer. **5**. A regulatory agency established by the US Congress to administer the nation's environmental laws. Also called the US EPA.

Error: Discrepancy between a computed, observed, or measured value or condition and the true specified or theoretically correct value or condition.

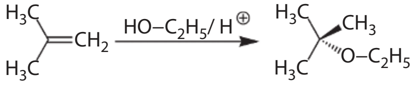
Ethane (C_2H_6): A colorless gas; a minor constituent of natural gas and a component in refinery gas that, along with methane, is typically used as refinery fuel. An important feedstock for making ethylene.

Ether ($C_2H_5OC_2H_5$): **1**. A generic term applied to a group of organic chemical compounds composed of carbon, hydrogen, and oxygen, characterized by an oxygen atom attached to two carbon atoms (e.g., methyl tertiary butyl ether). **2**. Any carbon compound containing the functional group (C–O–C). A commonly used ether is diethyl ether, which is used as an anesthetic.



Ethylene (C_2H_4): A colorless gas created by cracking processes. In refineries, it is typically burned with the methane and ethane. In chemical plants, it is purposefully made in ethylene plants and it is a basic building block for a wide range of products including polyethylene and ethyl alcohol.

ETBE: Ethyl Tertiary Butyl Ether $(CH_3)_3$ CO C_2H_5: **1**. A colorless, flammable, oxygenated hydrocarbon blend stock. It is produced by the catalytic etherification of ethanol (C_2H_5OH) with isobutylene ($_iC_4H_8$). **2**. An oxygenated gasoline blending compound to improve octane and reduce carbon monoxide emissions. It is commonly used as an oxygenate gasoline additive in the production of gasoline from crude oil.



Equation of state: A relationship that links the pressure, volume, and temperature of an amount of a substance. It is used to determine thermodynamic properties such as liquid and vapor densities, vapor pressures, fugacities, and deviations from ideality and enthalpies. Various equations of state have been developed to predict the properties of real substances. Commonly used equations of state include the ideal gas law; virial equation; van der Waals' equation; and Peng-Robinson, Soave–Redlich Kwong, and Lee–Kesler equations. Cubic equations are relatively easy to use and are fitted to experimental data. The van der Waals equation is comparatively poor at predicting state properties. The Lee–Kesler model, which is based on the theory of corresponding states, uses reduced temperature and pressure and covers a wide range of temperatures and pressures (see Table 1).

Equilibrium: A condition or state in which a balance exists within a system, which may be physical or chemical. A system is in equilibrium if it shows no tendency to change its properties with time. Static equilibrium occurs if there is no transfer of energy across the system boundary, whereas dynamic equilibrium is when a transfer occurs, but the net effect of the energy is zero. Thermodynamic equilibrium occurs when there is no heat or work exchange between a body and its surroundings. Chemical equilibrium occurs when a chemical reaction takes place in the forward direction, when reactants form products at exactly the same rate as the reverse reaction of products reverts to their original reactant form.

Equilibrium constant (K_c): A reversible process, chemical or physical, in a closed system will eventually reach a state of equilibrium. The equilibrium is dynamic and may be considered as a state at which the rate of the process in one direction exactly balances the rate in the opposite direction. For a chemical reaction, the equilibrium concentrations of the reactants and products will remain constant provided that the conditions remain unchanged for the homogeneous system:

$$wA + xB \leftrightarrow yC + zD$$

The ratio of the molar concentrations of products to reactants remain constant at a fixed temperature; the equilibrium constant, K_c, is:

$$K_c = \frac{[C]^y [D]^z}{[A]^w \, [B]^x}$$

Table 1 Useful equations of state.

Name	Equation	Equation constants and functions
(1) ideal gas law	$P = \frac{RT}{V}$	None
(2) Generalized	$P = \frac{ZRT}{V}$	$Z = Z(p_r, T_r, Z_c$ or $\omega)$ as derived from data
(3) van-der-Waals	$P = \frac{RT}{(V-b)} - \frac{a}{V^2}$	a and b are species-dependent constants and estimated from the critical temperature and pressure
(4) Redlich-Kwong (R-K)	$P = \frac{RT}{(V-b)} - \frac{a}{V^2 + bV}$	$b = 0.08664\, RT_c/P_c$ $a = 0.42748\, R^2\, T_c^{2.5}/P_c\, T^{0.5}$
(5) Soave-Redlich-Kwong (S-R-K or R-K-S)	$P = \frac{RT}{(V-b)} - \frac{a}{V^2 + bV}$	$b = 0.08664\, RT_c/P_c$ $a = 0.42748\, R^2\, T_c^2[1 + f_\omega(1 - T_r^{0.5})]^2/P_c$ $f_\omega = 0.48 + 1.574\omega - 0.176\omega^2$
(6) Peng-Robinson (P-R)	$P = \frac{RT}{V-b} - \frac{a}{V^2 + 2\,bV - b^2}$	$b = 0.07780R\, T_c/P_c$ $a = 0.45724\, R^2\, T_c^2[1 + f_\omega(1 - T_r^{0.5})]^2/P_c$ $f_\omega = 0.37464 + 1.54226\omega - 0.26992\omega^2$

For the Haber process for the synthesis of ammonia, nitrogen is reacted with hydrogen as:

$$N_2\,(g) + 3H_2\,(g) \leftrightarrow 2NH_3\,(g)$$

The equilibrium constant is expressed as partial pressure:

$$K_c = \frac{[NH_3]^2}{[N_2]\,[H_2]^3} = \frac{p^2_{NH_3}}{p_{N_2}\,p_{H_2^3}}$$

Equilibrium K-value (K-value): The ratio of the mole fraction in the vapor divided by the mole fraction in the liquid for a component in the equilibrium state. Each K-value corresponds to a given temperature, pressure, and mixture composition.

Equilibrium ratio (K): the ratio of the mole fraction in the vapor phase of a component in a mixture, y, to the mole fraction in the liquid phase, x, at equilibrium:

$$K_A = \frac{y_A}{x_A}$$

It is a function of both temperature and pressure. The relative volatility, α, is less dependent on temperature and pressure than the equilibrium constant where for an ideal mixture of two components, A and B:

$$\alpha_{AB} = \frac{K_A}{K_B}$$

Equilibrium-Flash Vaporizer: When a mixture is heated without allowing the vapor to separate from the remaining liquid, the vapor assists in causing the high-boiling parts of the mixture to vaporize. Thus continuous-flash vaporization is used in almost all plant operations. The equipment is used to determine a flash vaporization curve, where a series of runs at different temperatures are conducted, and each run constitutes one point (of temperature and percentage vaporized) on the flash curve.

Equipment Reliability: The probability that, when operating under stated environment conditions, the process equipment will perform its intended function adequately for a specified exposure period.

Equivalent Length: A method used to determine the pressure drop across pipe fittings such as valves, bends, elbows, and T-pieces. The equivalent length of a fitting is the length of a pipe that would give the same pressure drop as the fitting. Since each size of a pipe or fitting requires a different equivalent length for any particular type of fitting, it is usual to express the equivalent length as so many pipe diameters and this number is independent of the pipe. For example, if a valve in a pipe of diameter, d, is said to have an equivalent length, n, of pipe diameters, then the pressure drop due to the valve is the same as that offered by a length, and of the pipe.

Ergun Equation: Sabri Ergun developed an equation in 1952 to determine the pressure drop per unit length of a fixed bed of particles such as a catalyst at incipient gas velocity, v:

$$\frac{-\Delta p}{L} = \frac{150(1-e)^2\,\mu v}{\phi e^3 d^2} + \frac{1.75(1-e)\rho v^2}{\phi e^3 d}$$

where $-\Delta p/L$ is the pressure drop over the depth of bed, e is the bed voidage, d is the mean particle diameter, ρ is the fluid density, μ is the fluid viscosity, and ϕ is the sphericity. The incipient point of fluidization corresponds with the highest pressure drop at the minimum fluidization velocity.

Erosion: The physical removal of a material from a surface by mechanisms that exclude chemical attack. The usual phenomenon that causes erosion is impingement by either liquid droplets or entrained solid particles. If there are no corrosive substances present, then, in many cases, the most common mechanism for material damage due to erosion is impingement by solid particles.

Exothermic reaction: **1**. A chemical reaction that gives out/liberates heat. No energy input is required for the reaction to proceed. It has a negative enthalpy change, and therefore, under the appropriate conditions, the reaction will occur spontaneously. Chemical reactors used to contain exothermic reactions, therefore, require cooling facilities to remove the excess heat that is generated and to maintain a constant temperature. **2**. A reaction in which heat is evolved. Alkylation, polymerization, and hydrogenation reactions are exothermic reactions.

Expansion Loop: Piping thermally expands as it gets hot. Allowance must be made for the growth in pipe length; otherwise, the pipe will break by cracking at its welds. A fractionator at the Good Hope Refinery in the United States was burned down because of such an omission.

Explosion: **1**. The sudden conversion of potential energy (chemical or mechanical) to kinetic energy with the production and release of gases under pressure or the release of gas under pressure. **2**. A release of energy that causes a pressure discontinuity or blast wave.

Exports: Shipments of crude oil and petroleum products from countries, e.g., in the United States—shipments from the 50 states and the District of Columbia to foreign countries, Puerto Rico, the Virgin Islands, and other US possessions and territories.

Failure: **1**. Termination of the ability of a functional unit to perform a required function. **2.** An unacceptable difference between the expected and observed performance.

Fail Safe: A system design or condition such that the failure of a component, subsystem, or system or input to it will automatically revert to a predetermined safe static condition or state of least critical consequence for the component, subsystem, or system.

Fail Steady: A condition wherein the component stays in its last position when the actuating energy source fails. May also be called Fail in Place.

Failure Mode: The action of a device or system to revert to a specified state upon the failure of the utility power source that normally activates or controls the device or system. Failure modes are normally specified as fail open (FO), fail close (FC), or fail steady (FS), which will result in a fail-safe or fail-to-danger arrangement.

Failure Mode and Effects Analysis (FMEA): A systematic, tabular method for evaluating and documenting the causes and effects of known types of component failures.

Fault: Abnormal condition that may cause a reduction in, or loss of, the capability of a functional unit to perform a required function.

Fault Tree: A logic method that graphically portrays the combinations of failures that can lead to a specific main failure or accident of interest.

Field Production: Represents crude oil production on leases, natural gas liquids production at natural gas processing plants, a new supply of other hydrocarbons/oxygenates and motor gasoline blending components, and fuel ethanol blended into finished motor gasoline.

Final Boiling Point (FBP): The final boiling point of a cut, usually on an ASTM distillation basis.

Fifteen-five (15/5) distillation: A laboratory batch distillation performed in a 15-theoretical plate fractionating column with a 5:1 reflux ratio. A good fractionation results in accurate boiling temperatures. For this reason, this distillation is referred to as the true boiling point distillation. This distillation corresponds very closely to the type of fractionation obtained in a refinery.

Fire:**1**. A combustible vapor or gas combined with an oxidizer in a combustion process manifested by the evolution of light, heat, and flame. **2**. The rapid thermal oxidation (combustion) of a fuel source, resulting in heat and light emission. There are various types of fire, classified by the type of fuel and associated hazards. In the United States, the National Fire Protection Association (NFPA) classifies fires and hazards by the types of fuels or combustibles in order to facilitate the control and extinguishing of fires:

Class A	Ordinary combustibles such as wood, cloth, paper, rubber, and certain plastics
Class B	Flammable or combustible liquids, flammable gases, greases, and similar materials
Class C	Energized electrical equipment
Class D	Combustible metals, such as magnesium, titanium, zirconium, sodium, and potassium

Fireball: The atmosphere burning of a fuel–air cloud in which the energy is mostly emitted in the form of radiant heat. The inner core of the fuel release consists of almost pure fuel, whereas the outer layer in which ignition first occurs is a flammable fuel–air mixture. As the buoyancy forces of the hot gases begin to dominate, the burning cloud rises and becomes more spherical in shape.

Fire Point: Is the temperature well above the flashpoint where the product could catch a fire. The fire point and flash point are always taken care of in the day-to-day operation of a refinery. (See also Flashpoint).

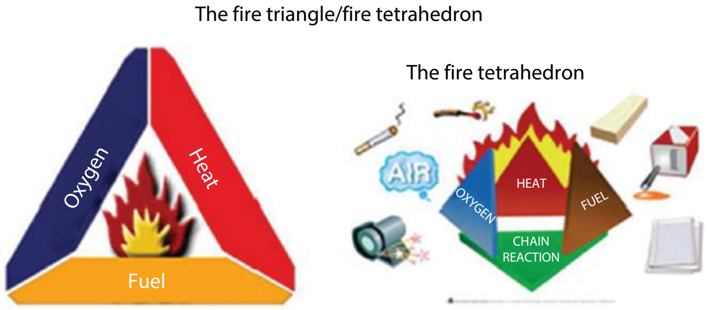
Fireproof: Resistant to specific fire exposure. Essentially nothing is absolutely fireproof, but some materials or building assemblies are resistant to damage or fire penetration at certain levels of fire exposures that may develop in the petroleum, chemical, or related industries.

Fireproofing: A common industry term used to denote the materials or methods of construction used to provide fire resistance for a defined fire exposure and specified time. Essentially nothing is fireproof if it is exposed to high temperatures for an extended period of time.

Fire Suppression System: A method, device, or system used to detect a fire or ignition source and to extinguish the fire in sufficient time so as to prevent structural damage and/or the debilitation of personnel.

Fire triangle: A way of illustrating the three factors necessary for the process of combustion, which are fuel, oxygen, and heat. All three are required for combustion to occur. A fire can therefore be prevented or extinguished by removing one of the factors. A fire is not able to occur without sufficient amounts of all three (See Figure 11).

First Law of Thermodynamics: A law that is applied to the conservation of energy in which the change in internal energy, ΔU, of a system is equal to the difference in the heat added, Q, to the system and the work done by the system:



Fire safety, at its most basic, is based upon the principle of keeping fuel sources and ignition source separate.

Figure 11 Diagram of a fire triangle.

$$\Delta U = Q - W$$

When considering chemical reactions and processes, it is more usual to consider situations where work is done on the system rather than by it.

Fittings: Connections and couplings used in pipework and tubing. The type of fittings used depends largely on the wall thickness as well as in part on the properties of the pipes and tubes including welds, flanges, and screw fittings. Fittings include elbow, bends, tees, reducers, and branches.

Fixed Bed: A place in a vessel for a catalyst through or by which feed can be passed for reactions, as opposed to a fluid bed, where the catalyst moves with the feed.

Fixed Bed Reactor: A reactor in which the catalyst is loaded into an immovable bed. The reactants enter the top of the bed, and the products exit from the bottom of the bed. The process must be taken offline, and hot gases are circulated through the catalyst bed to burn off coke deposits and restore the catalyst activity (See Figure 12).

Flame: The glowing gaseous part of a fire.

Flammable: **1**. A substance or material that has the ability to support combustion and be capable of burning with a flame. It is easily ignited or highly combustible. The term is more widely used than inflammable as this is often confused with incombustible, which means an inability or lack of ability to combust. A flammable liquid is a liquid that has the capability of catching a fire. In the United States, the National Fire Protection Association defines a flammable liquid as a liquid that has a flashpoint below 100°F (37.8°C) and a vapor pressure not exceeding 40 psia (2.72 bara) at that temperature. **2**. In general sense, refers to any material that is easily ignited and burns rapidly. It is synonymous with the term inflammable that is generally considered obsolete due to its prefix, which may be incorrectly misunderstood as not flammable (e.g., incomplete is not complete).

Flammable Liquid: **1**. As defined by NFPA 30, a liquid having a flash point below 100°F (37.8°C) and having a vapor pressure not exceeding 2068 mm Hg (40 psia) at 100°F (37.8°C) as determined under specific conditions. **2**. Any liquid having a flash point below 100°F (37.8°C), except any liquid mixture having one or more components with a flash point at or above the upper limit that makes up 99% or more of the total volume of the mixture. **3**. Liquid with a flash point below 100°F (37.8°C). At that temperature, vapors from the substance can be ignited by a flame, spark, or other sources of ignition.

Flammability Limit: **1**. The flammability limit of a fuel is the concentration of fuel (by volume) that must be present in the air for ignition to occur when an ignition source is present. **2**. The range of gas or vapor amounts in air that

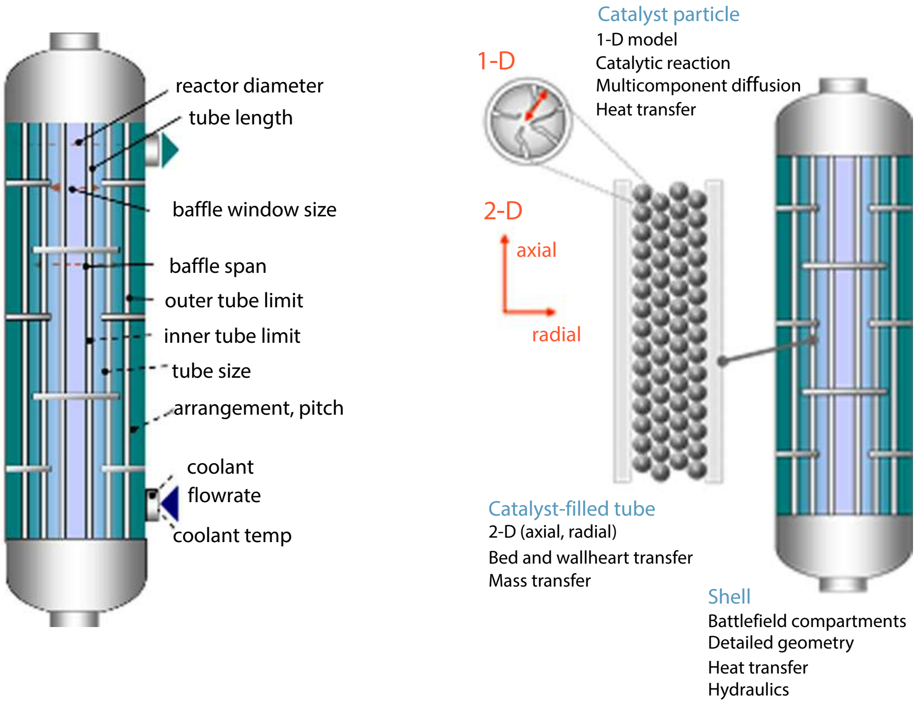


Figure 12 Diagrams of a fixed bed reactor.

will burn or explode if a flame or other ignition source is present. Importance: The range represents an unsafe gas or vapor mixture with air that may ignite or explode. Generally, the wider the range, the greater the fire potential.

Flange: It is a flat end of a pipe that is used to bolt up to a flange on another piece of piping. Bolts, with nuts at each end, are used to force the flanges together.

Flange Rating: Connections on vessels, spool pieces, and valves have a pressure rating called a flange rating. This rating can be confusing, e.g., a 150 psig flange rating is actually good for about 230 psi design.

Flare: **1**. A burner on a remote line used for the disposal of hydrocarbons during clean-up, for emergency, shutdowns, and the disposal of small-volume waste streams of mixed gases that cannot easily or safely be separated. **2**. A flame used to burn off unwanted natural gas; a "flare stack" is the steel structure on a processing facility from which gas is flared. **3**. An open flame used to burn off unwanted natural gas. **4**. To burn unwanted gas through a pipe or stack. **5**. The flame from a flare; the pipe or the stack itself.

Flared: Gas disposed of by burning in flares, usually at the production sites or at gas processing plants.

Flare Stack: The steel structure on an offshore rig or at a processing facility from which gas is flared.

Flare System: This is a piping network that runs through the plant to collect vents of gas so that they can be combusted at a safe location in the flare stack.

Flaring: Is the burning of a natural gas that cannot be processed or sold. Flaring disposes off the gas, and it releases emissions into the atmosphere.

Flaring/Venting: The controlled burning (flare) or release (vent) of natural gas that cannot be processed for sale or use because of technical or economic reasons.

Flashing: The vaporization of water or light ends as pressure is released during production or processing.

Flash Calculation: The determination of the compositions and quantities of liquid and vapor that co-exist in a mixture under equilibrium conditions.

Flash Chamber: A wide vessel in a vacuum flasher thermal cracking plant or similar operation into which a hot stream is introduced, causing the lighter fractions of that stream to vaporize and leave by the top.

Flash Fire: The combustion of flammable vapor and air mixture in which flame passes through that mixture at less than sonic velocity, such that negligible damaging overpressure is generated.

Flash Point: **1**. The minimum temperature at which a liquid, under specific test conditions, gives off sufficient flammable vapor to ignite momentarily on the application of an ignition source. **2**. The lowest temperature at which any combustible liquid will give off sufficient vapor to form an inflammable mixture with air (i.e., that can be readily ignited). Flash points are used to specify the volatility of fuel oils, mostly for safety reasons. They are generally an indication of the fire and explosion potential of a product; ASTM D-56, D-92, D-93, D-134, and D-1310. **3**. Hold a flame over a cup of diesel fuel; it will start to burn at its 160°F (71°C) flash temperature. Gasoline's flash point is below room temperature. For jet fuel, it is 110°F (43°C). The lighter the hydrocarbon, the lower the flash point.

Flash Tank: A container where the separation of gas and liquid phases is achieved after pressure reduction in flow fluid. Both phases appear when pressure is decreased as a consequence of the Joule–Thompson effect.

Flash Vapors: Vapors released from a stream of natural gas liquids as a result of an increase in temperature or a decrease in pressure.

Flash Zone: The section of a distillation/fractionating column containing the column feed nozzle(s). The column feed separates or "flashes" into liquid and vapor as it expands through the feed nozzle(s) and enters the column.

Flexicoking: A thermal cracking process that converts heavy hydrocarbons such as crude oil, oil sands bitumen, and distillation residues into light hydrocarbons. Feedstocks can be any pumpable hydrocarbons including those containing high concentrations of sulfur and metals.

Flooding: **1**. An excessive buildup of liquid in absorption columns or on the plate of a distillation column. It is due to high vapor flow rates up the column. In distillation columns, this is caused by high heating rates in the reboiler. **2**. An all-inclusive term that is given to non-equilibrium behavior in a distillation/fractionating column because of larger flows of liquid and/or vapor than the column can process. Flooding can be caused by liquid backing up in the column, vapor blowing through the column and lifting the liquid off the trays, etc. All columns are designed to handle about 80% of the flow before flooding occurs. Sometimes, flooding is caused by mechanical restrictions or damage to the internals in the column. The ***flooding point*** is a condition in a packed column such as an absorption column that receives a countercurrent flow of gas at the bottom and a liquid descending under gravity from the top where there is an insufficient liquid holdup in the packing for mass transfer to take place effectively. The liquid, therefore, descends to the bottom of the column without mass transfer. The rate of flow through the packing for effective mass transfer is controlled by the pressure drop across the packing material.

Flow: The movement of fluid under the influence of an external force such as gravity or a pump.

Flowline: A pipeline that carries materials from one place to another. In the offshore industry, a flowline is a pipeline that carries oil on the seabed from a well to a riser. On a process flow diagram, the flowline is indicated by a line entering and leaving a vessel or unit operation. An arrow indicates the direction of flow.

Flow Meter: A device used to measure the flow of process fluids. Flow meters are mainly classified into those that are intrusive and those that are non-intrusive to the flow of the fluid. Flow meters include differential pressure meters; positive displacement meters; and mechanical, acoustic, and electrically heated meters. The measurement of the flow of process fluids is essential not only for safe plant control but also for fiscal monitoring purposes. It is essential to select correctly the flow meter for a particular application, which requires a knowledge and comprehension of the nature of the fluid to be measured and an understanding of the operating principles of flow meters.

Flow Rate: The movement of material per unit time. The material may be gas, liquid, or solid particulates in suspension or a combination of all of these and expressed on a mass, volumetric, or molar basis. The volumetric flow of material moving through a pipe is the product of its average velocity and the cross-sectional area of the pipe.

Flow Regime: The behavior of a combined gas and liquid flow through a duct, channel, or pipe can take many forms, and there are descriptions used to define the possible flow patterns. Depending on the conditions of the flow of the two phases, one phase is considered to be the continuous phase, while the other is the discontinuous phase. An example is the flow of mist or fine dispersion of liquid droplets in a gas phase. The smaller the liquid droplets, the higher the surface tension effects. A distortion of the discontinuous phase results in the shape becoming non-spherical. Also, there is a tendency for the liquid phase to wet the wall of the pipe and for the gas phase to congregate at the center. An exception to this is in evaporation such as in refrigeration where nuclear boiling occurs on the pipe surface, resulting in a vapor film or bubbles forming at the surface with a central core of liquid. The flow of fluids through pipes and over surfaces can be described as:

1.	Steady flow in which the flow parameters at any given point do not vary with time.
2.	Unsteady flow in which the flow parameters at any given point vary with time.
3.	Laminar flow in which the flow is generally considered to be smooth and streamlined.
4.	Turbulent flow in which the flow is broken up into eddies and turbulence.
5.	Transition flow, which is a condition lying between the laminar and turbulent flow regimes.

Flow regime maps are charts representing the various flow patterns that are possible for two-phase gas–liquid flow in both horizontal and vertical pipes and tubes. There are many types of flow regime maps that have been developed. The simplest form of the map involves a plot of superficial velocities or flow rates for the two phases, with the most widely used generalized flow regime map for horizontal flow as shown below (See Figure 13).

The maps are populated with experimental data in which lines are drawn to represent the boundaries between the various flow regimes. These include dispersed, bubble or froth, wavy, annular, stratified, slug, and plug flow. The boundaries between the various flow patterns are due to the regime becoming unstable as it approaches the boundary with the transition to another flow pattern. As with the transition between laminar and turbulent flow in a pipe, the transitions in a flow regime are unpredictable. The boundaries are therefore not distinct lines but loosely defined transition zones. A limitation of the maps is that they tend to be specific to a particular fluid and pipe. The seven types of flow regimes in the order of increasing gas rate at constant liquid flow rate are given below (See Figure 13).

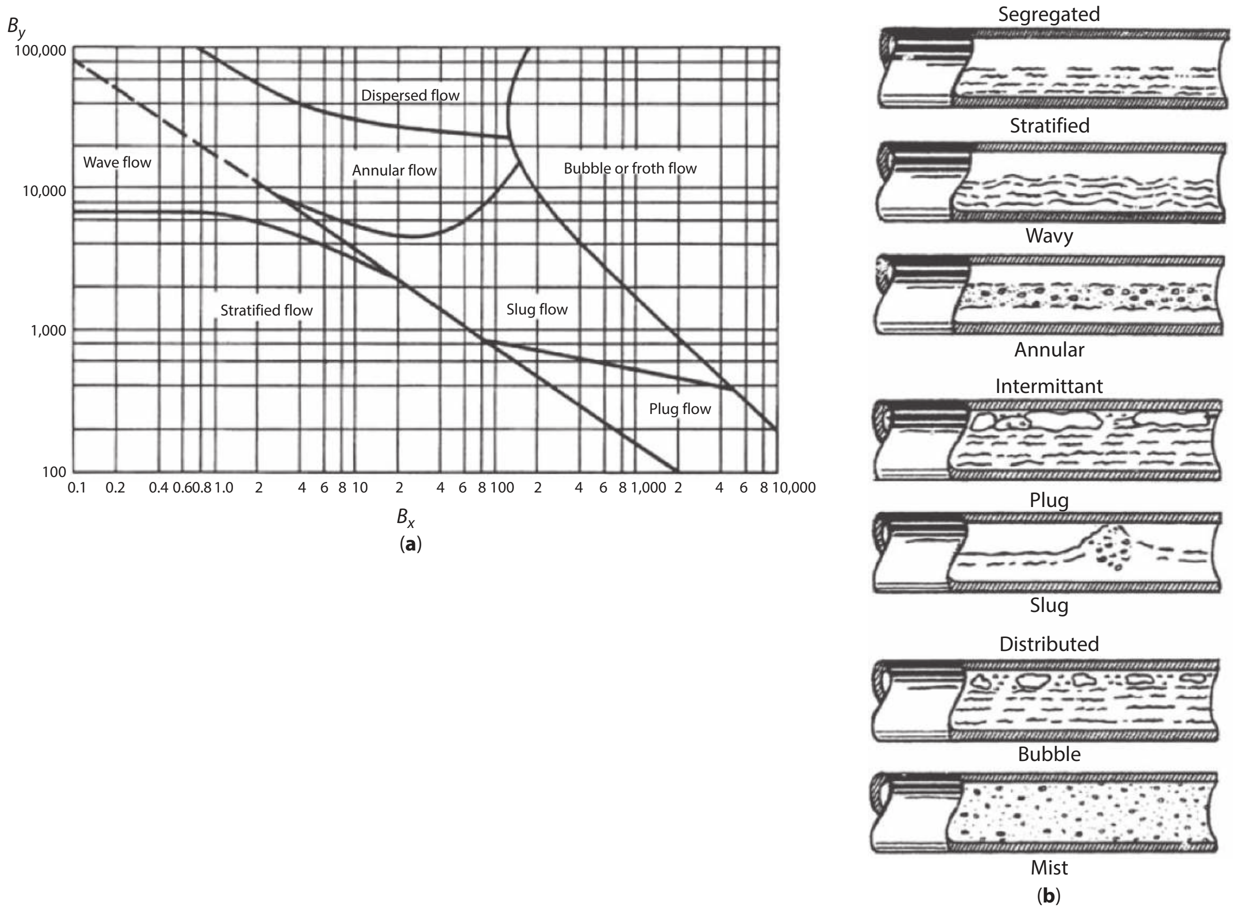


Figure 13 (a) Flow patterns for horizontal two-phase flow (Based on data from 1, 2, and 4 in. pipe by Baker, O., Oil & Gas J., Nov. 10, p. 156, 1958). (b) Representative forms of horizontal two-phase flow patterns as indicated in Figure 10a.

Bubble or Froth Flow: Bubbles of gas are dispersed throughout the liquid and are characterized by bubbles of gas moving along the upper part of the pipe at approximately the same velocity as the liquid. This type of flow can be expected when the gas content is less than about 30% of the total (weight) volume flow rate. (Note: About 30% gas by weight is over 99.9% by volume, normally.)

Plug Flow: Alternate plugs of liquid and gas move along the upper part of the pipe, and liquid moves along the bottom of the pipe.

Stratified Flow: Liquid flows along the bottom of the pipe, while gas flows over a smooth liquid–gas interface.

Wave Flow: Wave flow is similar to stratified flow, except that the gas is moving at a higher velocity and the gas–liquid interface is distributed by waves moving in the direction of flow.

Slug Flow: This pattern occurs when waves are picked up periodically by the more rapidly moving gas. These form frothy slugs that move along the pipeline at a much higher velocity than the average liquid velocity. This type of flow causes severe and, in most cases, dangerous vibrations in equipment because of the high-velocity slugs against fittings.

Annular Flow: In annular flow, liquid forms around the inside wall of the pipe and gas flows at a high velocity through the central core.

Dispersed, Spray, or Mist flow: Here, all of the liquid is entrained as fine droplets by the gas phase. This type of flow can be expected when the gas content is more than roughly 30% of the total weight flow rate. Some overhead condenser and reboiler-return lines have dispersed flow.

Flowsheet: A schematic diagram or representation of a process illustrating the layout of process units and their functions linked together by interconnecting process streams. The development of a flowsheet involves the process synthesis, analysis, and optimization. The heat and material balances are solved using thermodynamic properties and models. An economic analysis is also completed as well as a safety and environmental impact assessment. The choice of equipment and their interconnectivity are optimized along with the choice of operating parameters such as temperature, pressure, and flows. Steady-state flowsheet computer software packages are frequently used to develop flowsheets.

Flue: Passage through which flue gases pass from a combustion chamber to the outside atmosphere.

Flue Gas: **1**. A mixture of gases produced as a result of combustion that emerges from a stack or chimney. The gases contain smoke, particulates, carbon dioxide, water vapor, unburnt oxygen, nitrogen, etc. An Orsat analysis is a reliable device to determine the composition of the flue gas and the efficiency of combustion, although it has been replaced by other techniques. **2**. Gas from the various furnaces going up to the flue (stack).

Fluid Catalytic Cracking (FCC): A thermal process in which the oil is cracked in the presence of a finely divided catalyst that is maintained in an aerated or fluidized state by the oil vapors. The powder or fluid catalyst is continuously circulated between the reactor and the regenerator, using air, oil vapor, and steam as the conveying media. The most commonly used catalytic cracking process. The catalyst is a fine powder that is designed to form a fluidized bed in the reactor and regenerator (Figure 14).

Fluid Coking: **1**. A coking process in which the feed is preheated and sprayed into a reactor where it contacts a hot fluidized bed of coke returning from a burner vessel. The hot oil products are stripped from the coke, which is circulated back to the burner vessel. Coke not returned to the reactor from the burner vessel is withdrawn as a coke product. **2**. A thermal cracking process utilizing the fluidized solids technique to remove carbon (coke) for the continuous conversion of heavy, low-grade oils into lighter products.

Fossil Fuels: Fuels formed by natural processes such as anaerobic decomposition of buried dead organisms. The age of the organisms and their resulting fossil fuels is typically millions of years and sometimes exceeds 650 million years. Fossil fuels contain a high percentage of carbon and include coal, petroleum, and natural gas. Other more commonly used derivatives of fossil fuels are kerosene and propane. They range from volatile materials with low carbon-to-hydrogen ratios like methane to liquid petroleum, to non-volatile materials composed of almost pure carbon, like anthracite coal. Methane can be found in hydrocarbon fields, alone, associated with oil, or in the form of methane clathrates (See Figure 15).

Georg Agricola, in 1556, first introduced the theory that fossil fuels were formed from the fossilized remains of dead plants by exposure to heat and pressure in the earth's crust over millions of years and later expounded by Mikhail Lomonosov in the 18th century. Coal is one of the fossil fuels (See Figure 15).

The use of fossil fuels raises serious environmental concerns. The burning of fossil fuels produces around 21.3 billion tonnes (21.3 gigatonnes) of carbon dioxide (CO_2) per year, but it is estimated that natural processes can only absorb about half of that amount, so there is a net increase of 10.65 billion tonnes of atmospheric carbon dioxide per year (one tonne of atmospheric carbon is equivalent to 44/12 or 3.7 tonnes of carbon dioxide). Carbon dioxide is one of the

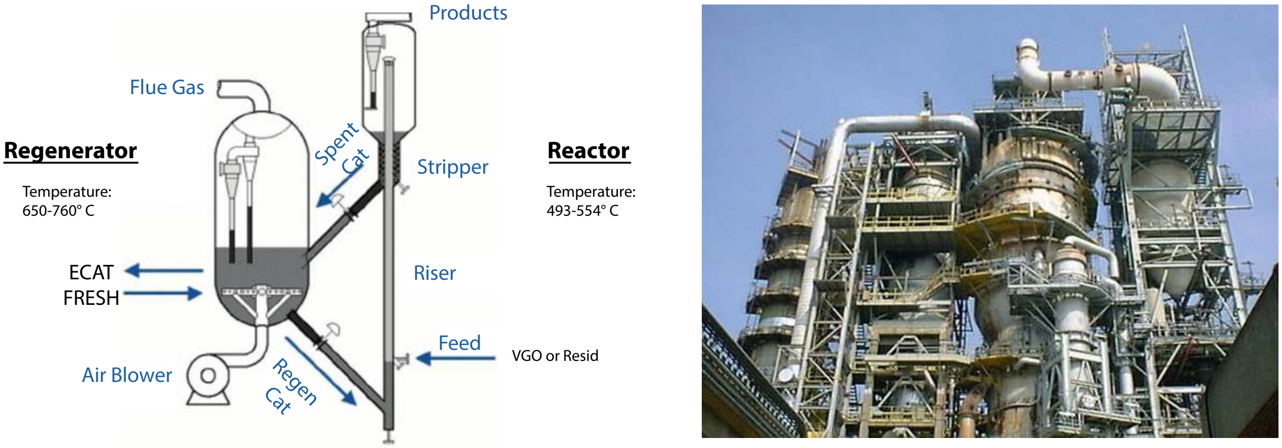


Figure 14 Fluid catalytic cracking unit.

Figure 15 Coal.

greenhouse gases that enhances the radiative forcing and contributes to global warming, causing the average surface temperature of the earth to rise with major adverse climatic effects. A global movement toward the generation of renewable energy is therefore essential to help reduce global greenhouse gas emissions.

The ratio of gross domestic product to kilograms of fossil fuel carbon consumed, for the world's 20 largest economies. The two countries with the highest GDP per kilogram carbon ratios, Brazil and France, produce large amounts of hydroelectric and nuclear power, respectively (See Figure 16).

Fractionation: The general name given to a process for separating mixtures of hydrocarbons or other chemicals into separate streams or cuts or fractions.

Free Carbon: The organic materials in tars that are insoluble in carbon disulfide.

Free Energy of Formation: The change in free energy when a compound is formed from its elements with each substance in its standard state at 77°F (25°C). The heat of reaction at 25°C may be calculated by subtracting the sum of the free energies of formation of the reactants from the sum of the free energies of formation of the products.

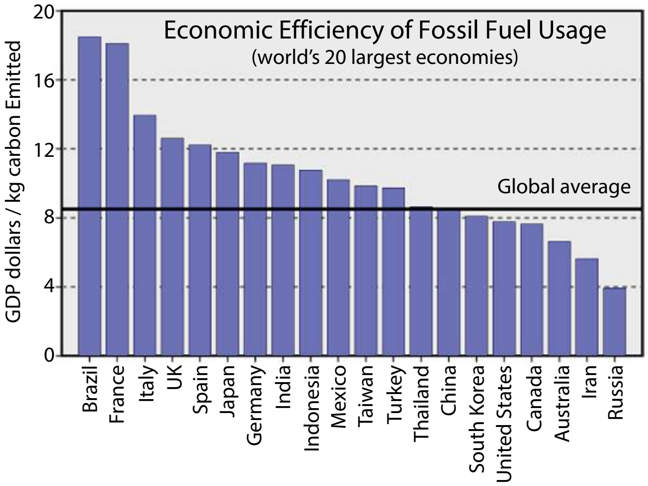


Figure 16 Economic efficiency of fossil fuel usage.

Free Water: Condensed water that exits as a separate liquid phase. Most refinery distillation columns are designed such that free water will not be present, since it can result in column upsets and promote the corrosion of the metal in the column.

Freeze Point: **1**. The temperature at which the hydrocarbon liquid solidifies at atmospheric pressure. It is an important property for kerosene and jet fuels because of the very low temperatures encountered at high altitudes in jet planes. One of the standard test methods for the freeze point is ASTM D4790. **2**. The temperature at which a chilled petroleum product becomes solid and will no longer pour when a sample tube is tipped. Freeze point is a laboratory test. **3**. The temperature at which crystals first appear as a liquid is cooled, which is especially important in aviation fuels, diesel, and furnace oil.

Front End Engineering and Design (FEED) Contract: **1**. A legal agreement setting out the terms for all activities required to define the design of a facility to a level of definition necessary for the starting point of an EPC contract. **2**. Generally, the second contracting phase for the development of the export facilities in the LNG chain provides a greater definition than the prior conceptual design phase. In an LNG project, the single most important function of an FEED contract is to provide the maximum possible definition for the work to be ultimately performed by the engineering procurement and construction (EPC) contractor. **3**. A study used to analyze the various technical options for new field developments with the objective to define the facilities required. **4**. The stage of design between concept evaluation and detailed design during which the chosen concept is developed such that most key decisions can be taken. The output of FEED includes an estimate of the total installed cost and schedule. *See also Engineering, Procurement, and Construction Contract.*

Fuel Gas: **1**. A process stream internal to a facility is used to provide energy for operating the facility. **2**. Gas used as fuel in a liquefaction plant. It typically involves processing waste streams to LNG that are not profitable. It is used in gas turbines, boilers, and reaction furnaces.

Fuel Oil: Usually residual fuel but sometimes distillate fuel.

Fuel Oil Equivalent (FOE): The heating value of a standard barrel of fuel oil, equal to 6.05 × 10^6 Btu (LHV). On a yield chart, dry gas and refinery fuel gas are usually expressed in FOE barrels.

Furnace Oil: A distillate fuel made of cracked and straight-run light gas oils primarily for domestic heating because of its ease of handling and storing.

FVT: The final vapor temperature of a cut. The boiling ranges expressed in this manner are usually on a crude assay, true boiling point basis.

Gas/Liquid Chromatography (GC, GLC): Equipment used to determine the composition of a sample in the laboratory.

Gap: Gas is usually based on ASTM 86 distillation temperatures and is defined as the 5% distilled temperature of a distillation column product minus the 95% distilled temperature of the next higher product in the column. When the difference is positive, the difference is called a gap. When the difference is negative, the difference is sometimes called an overlap. The gap or overlap is a measure of the sharpness of the separation between adjacent products in a distillation column.

Gas Cap: An accumulation of natural gas at the top of a crude oil reservoir. The gas cap often provides the pressure to rapidly evacuate the crude oil from the reservoir.

Gasket: This is the softer material that is pressed between flanges to keep the fluid from leaking. Using the wrong gasket is a common cause of fires in process plants. Gaskets have different temperature and pressure ratings.

Gas Oil: **1**. Any distillate stream having molecular weights and boiling points higher than heavy naphtha (>400°F or 205°C). The name gas oil probably traces its roots to "gasoline" bearing oil in the early days of refining. Early refiners used thermal cracking processes to produce more motor gasoline (MOG) from gas oil stocks. **2**. The term is used for petroleum stocks with boiling ranges between approximately 650–1100°F (344–594°C). Unreacted gas oils are produced by distilling crude oil in crude and vacuum columns. Cracked gas oils are produced in refinery reaction processes, such as thermal and catalytic cracking, coking, visbreaking, and hydrocracking.

Gasoline: **1.** A light petroleum product in the range of 80–400°F (27–204°C) for use in spark-ignition internal combustion engines. **2**. An all-inclusive name for petroleum stocks that are used as fuel for internal combustion engines. Retail gasoline is a blend of several refinery gasolines and must meet the specifications of octane, Reid vapor pressure, distilling boiling range, sulfur content, and so on. Additives such as ethers or alcohols are used to improve the octane for the blended product.

Gasoline Blending Components: Naphthas that will be used for blending or compounding into finished aviation or motor gasoline (e.g., straight-run gasoline, alkylate, reformate benzene, toluene, and xylenes). Excludes oxygenates (alcohols, ethers), butane, and natural gasoline.

Gas sweetening: A process used to remove hydrogen sulfide and mercaptans from natural gas. Commonly used in petroleum refineries, the gas treatment uses amine solutions such as monoethanolamine. The process uses an absorber unit and regenerator. The amine solution flows down the scrubber and absorbs the hydrogen sulfide as well as carbon dioxide from the upflowing gases. The regenerator is used to strip the amine solution of the gases for reuse. It is known as gas sweetening as the foul smell is removed from the gas.

Gas Treating: The amine treating of light gases to remove such impurities as H_2S and CO_2. Molecular sieves are also used to concentrate hydrogen streams by removing inerts and light hydrocarbon contaminants.

Gas Turbine: An engine that uses internal combustion to convert the chemical energy of a fuel into mechanical energy and electrical energy. It uses air, which is compressed by a rotary compressor driven by the turbine and fed into a combustion chamber where it is mixed with the fuel, such as kerosene. The air and fuel are burnt under constant pressure conditions. The combustion gases are expanded through the turbine, causing the blades on the shaft to rotate. This is then converted to electrical energy. Gas turbines are used in the process industries and on offshore gas platforms for electrical generation.

Global Warming Potential: The global warming potential (GWP) of a greenhouse gas is its ability to trap extra heat in the atmosphere over time relative to carbon dioxide (CO_2). The GWP depends on the following factors:

- The absorption of infrared radiation by a given gas.
- The spectral location of its absorbing wavelengths
- The atmospheric lifetime of the gas

Grain: A unit of mass where one pound is equivalent to 7000 grains and a specification of 0.25 grain of H_2S per 100 scf is equivalent to an H_2S concentration of 4.0 ppmv.

Gravity: The specific gravity (Sp.Gr.) of a stream, often expressed as API Gravity by petroleum refiners. The basis is always the density of water.

Gravity Curve: The gravity of the material distilled from a petroleum stock in a laboratory still. The gravity curve is plotted against the percent distilled for the stock. Gravity curves are most commonly reported for true boiling point distillation (Figure 17).

Greenhouse Gas (GHG): A greenhouse gas is a gas in an atmosphere that absorbs and emits radiation within the thermal infrared range. This process is the fundamental cause of the greenhouse effect. The primary greenhouse gases in the Earth's atmosphere are water vapor, carbon dioxide, methane, nitrous oxide, and ozone. Without greenhouse gases, the average temperature of the Earth's surface would be about 15°C (27°F) colder than the present average of 14°C (57°F).

Greenhouse Gas Emissions: An addition to the atmosphere of gases that are a cause of global warming, including carbon dioxide, methane, and others.

Greenhouse Gas Removal Technologies (or negative emissions): Methods that actively remove greenhouse gases from the atmosphere, ranging from engineering to nature-based solutions.

Gross Heating Value: Is the total energy transferred as heat in an ideal combustion reaction at a standard temperature and pressure in which all water formed appears as a liquid. The gross heating is an ideal gas property in a hypothetical state (the water cannot all condense to the liquid because some of the water would saturate the CO_2 in the products).

Gross Heating Value of Fuels (GHV): The heat produced by the complete oxidation of material at 60°F (25°C) to carbon dioxide and liquid water at 60°F (25°C).

Gross Input to Atmospheric Crude Oil Distillation Units: Total inputs to atmospheric crude oil distillation units. Include all crude oil, lease condensate, natural gas plant liquids, and unfinished oils. Liquefied refinery gases, slop oils, and other liquid hydrocarbons produced from oil sands, gilsonite, and oil shale.

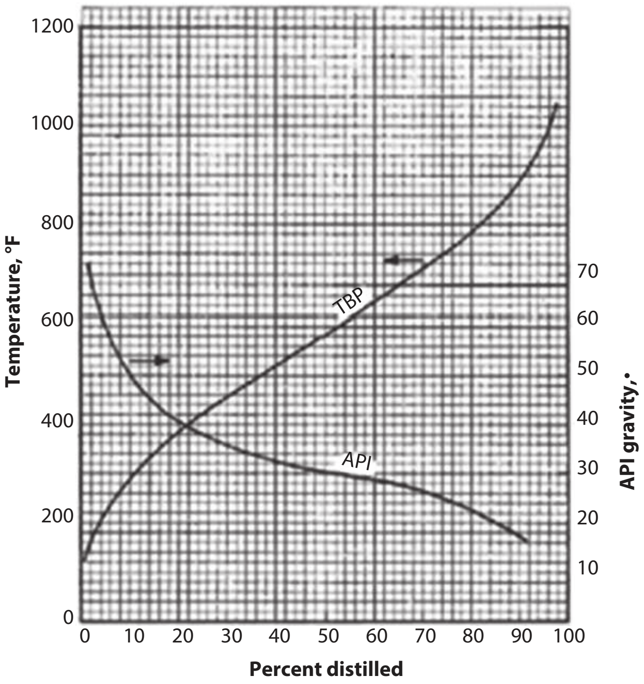


Figure 17 TBP and gravity—mid-percent curves.

Gum: A complex sticky substance that forms by the oxidation of gasoline, especially those stored over a long period of time. Gum fouls car engines, especially the fuel injection ports.

Harm: Physical injury or damage to the health of people, either directly or indirectly, as a result of the damage to property or to the environment.

Hazard: **1**. A condition or object that has the potential to cause harm. **2**. An unsafe condition, which, if not eliminated or controlled, may cause injury, illness, or death. **3**. A physical or chemical characteristic that has the potential for causing harm to people, the environment, or property. Examples of hazards:

- *Combustible/Flammable substance. E.g., Ethylene is flammable*
- *Corrosive. E.g., Sulfuric acid is extremely corrosive to the skin.*
- *Explosive substance. E.g., Acrylic acid can polymerize, releasing large amounts of heat.*
- *Toxic fumes. E.g., Chlorine is toxic by inhalation.*
- *Substances kept at high pressure in containment (e.g., a vessel, tank)*
- *Objects or material with a high or low temperature.*
- *Radiation from heat source.*
- *Ionizing radiation source.*
- *Energy release during the decomposition of a substance. E.g., Steam confined in a drum at 600 psig contains a significant amount of potential energy (Figure 18).*

Hazard Analysis: Is the first step in a process used to assess risk. The result of a hazard analysis is the identification of the different types of hazards.

It is assigned a classification, based on the worst-case severity of the end condition. Risk is the combination of probability and severity. Preliminary risk levels can be provided in the hazard analysis. The validation, more precise prediction (verification), and acceptance of risk is determined in the risk assessment (analysis). The main goal of both is to provide the best selection of the means of controlling or eliminating the risk.

Hazard Communication: Employees' "right-to-know" legislation requires the employers to inform employees (pretreatment inspectors) of the possible health effects resulting from contact with hazardous substances. At locations where this legislation is in force, employers must provide employees with information regarding any hazardous substances, which they might be exposed to under normal work conditions or reasonably foreseeable emergency conditions resulting from workplace conditions. OSHA's Hazard Communication Standard (HCS) (Title 29 CFR Part 1910.2100) is the Federal regulation and state statutes are called Worker Right-to- Know Laws.

Hazard Communication Program: A written plan to manage the hazards associated with the use of chemicals in the workplace.

Hazardous Chemical: A substance that may harm the worker either physically (e.g., fire, explosion) or chemically (e.g., toxic, corrosive).

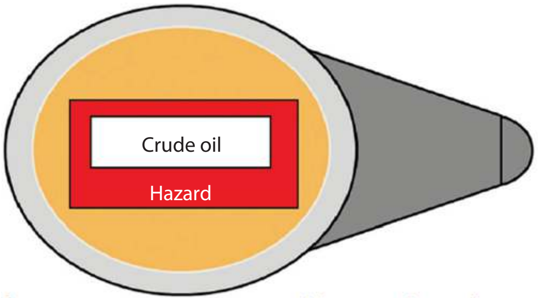
Hazardous Events: Hazardous event is defined as a hazardous situation that results in harm. Each identified hazard could give a number of different hazardous events. For each identified hazardous event, it should also be described which factors contribute to it.

For example, the hazardous combustible substance could give the following hazardous events:

- Pool fire outside a tank, due to leakage, when an ignition source is present.
- Flash fire inside a tank when an ignition source is present.

The factors that could contribute to the leakage in the tank, for instance, could be:

- *Bad connection joint.*
- *Gasket damage.*
- *Tube damage.*
- *Pipe damage.*



Consider pressure, temperature, composition, quantity, etc. into account.

Figure 18 A hazard.

Hazardous Situation: A circumstance in which a person is exposed to hazards.

HAZID/HAZOP: **1**. HAZard Identification / HAZard and Operability analysis systematic design review methods to identify and address hazards to ensure that the necessary safety measures to eliminate or mitigate hazards are incorporated in the design and operation of the unit. **2**. A qualitative process risk analysis tool used to identify hazards and evaluate if suitable protective arrangements are in place, if the process were not to perform as intended and unexpected consequences were to result.

HCGO: Heavy coker gas oil.

HCO: Heavy FCC cycle gas oil. See Heavy Cycle Oil.

Heart Cut Recycle: That unconverted portion of the catalytically cracked material is recycled to the catalytic cracker. This recycle is usually in the boiling range of the feed and, by definition, contains no bottoms. Recycling allows a less severed operation and suppresses the further cracking of desirable products.

Heat Balance: See Energy Balance.

Heat Exchanger: A pressure vessel for transferring heat from one liquid or vapor stream to another. A typical heat exchanger consists of a cylindrical vessel and nozzles through which one stream can flow and a set of pipes or tubes in series in the cylinder through which the other can flow. Heat transfer mechanisms are conduction and convection. See also Shell and Tube Heat Exchanger (See Figure 19).

Heat Flux: The rate of heat transfer per unit area normal to the direction of heat flow. It is the total heat transmitted by radiation, conduction, and convection.

Heat Pump: Thermodynamic heating/refrigerating system to transfer heat. The condenser and evaporator may change roles to transfer heat in either direction.

Heat Rate: The measure of efficiency in converting input fuel to electricity. Heat rate is expressed as the number of Btu of fuel (e.g., natural gas) per kilowatt-hour (Btu/kWh). The heat rate for power plants depends on the individual plant design, its operating conditions, and its level of electric power output. The lower the heat rate, the more efficient the plant.

Heat Recovery: Heat utilized that would otherwise be wasted from a heating system.

Heat Transfer Coefficient: Coefficient describing the total resistance to heat loss from a producing pipe to its surroundings. Includes heat loss by conduction, convection, and radiation.

Heating Oil: Any distillate or residual fuel. **1**. Oil used for residential heating. **2**. Trade term for the group of distillate fuel oils used in heating homes and buildings as distinguished from residual fuel oils used in heating and power installations. Both are burner fuel oils.

Heating Value: **1**. The average number of British thermal units per cubic foot of natural gas as determined from the tests of fuel samples. **2**. The amount of heat produced from the complete combustion of a unit quantity of fuel. **3**. The amount of energy or heat that is generated when a hydrocarbon is burned (chemically combined with oxygen). **4**. Energy released in the complete combustion of a unit of mass, matter, or volume of fuel in a stoichiometric mixture with air. **5**. The amount of heat produced by the complete combustion of a unit quantity of fuel.

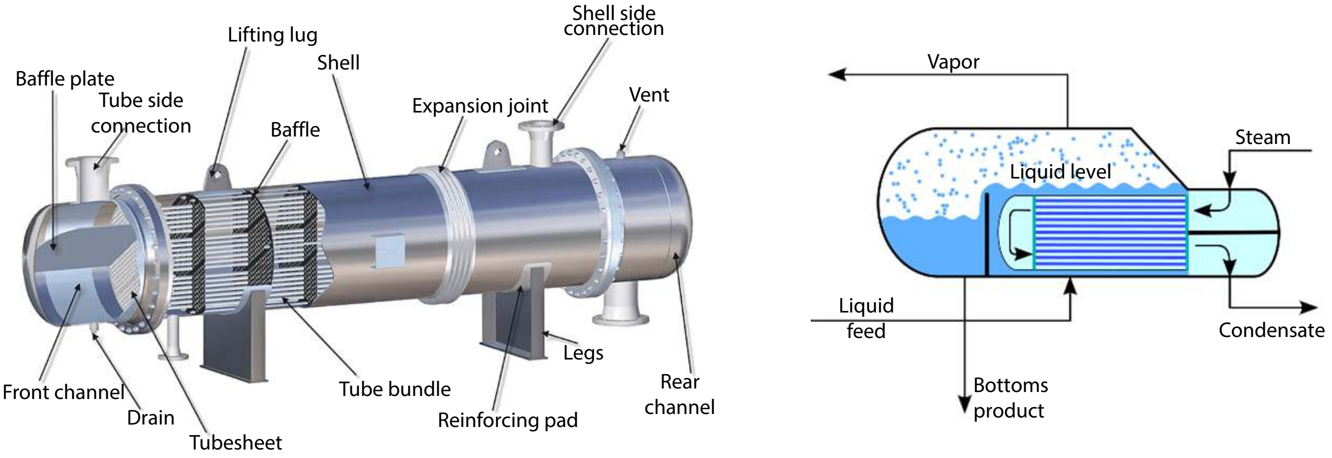


Figure 19 A shell and tube heat exchanger A reboiler.

Heat of Combustion: The amount of heat released in burning completely an amount of substance is its heat of combustion. The general formula for the combustion of a hydrocarbon compound is:

$$C_nH_{2n+2} + (3n+1)/2\ O_2 \rightarrow (n+1)\ H_2O + n\ CO_2 + \text{Energy}$$

Heat of Reaction: The heat release of heat absorbed when a chemical reaction takes place. The heat of reaction may be computed from the free energies of formation for the reacting components and the resultant products at the standard temperature of 77°F (25°C).

Heat of Vaporization: The amount of heat energy required to transform an amount of a substance from the liquid phase to the gas phase.

Heavy Crude: Crude oil of 20° API gravity or less; often very thick and viscous.

HCGO: Heavy coker gas oil.

Heavy Cycle Oil (HCO): Gas oil produced in an FCC operation that boils in the approximate TBP range of 400–1000°F (205–358°C). Heavy cycle oil is not generally withdrawn as a product, but it is recycled back to the reactor for further cracking to improve the overall conversion of the process.

Heavy Ends: The highest boiling portion of gasoline or other petroleum oil. The end point as determined by the distillate test reflects the amount and character of the heavy ends present in gasoline.

Heavy Gas Oil: Petroleum distillates with an approximate boiling range from 651 to 1000°F (344–538°C).

Heavy Key: A distributed component in a distillation section that is recovered in the (bottom) heavy product, with a small, specified amount leaving in the top product.

Heavy Oil: Lower gravity, often higher-viscosity oils. Normally less than 28° API gravity.

Hempel Distillation: The US Bureau of Mines' (now the Department of Energy, DOE) routine method of distillation. Results are frequently used interchangeably with TBP distillation.

Heptane (nC_7H_{16}): Normal heptane is a straight-chain alkane hydrocarbon with the chemical formula $H_3C(CH_2)_5CH_3$ or C_7H_{16}. Heptane (and its many isomers) is widely applied in laboratories as a totally non-polar solvent. As a liquid, it is ideal for transport and storage. Heptane is commercially available as mixed isomers for use in paints and coatings, as pure n- heptane for research and development and pharmaceutical manufacturing, and as a minor component of gasoline.

n-heptane is defined as the zero point of the octane rating scale. It is undesirable in gasoline because it burns explosively, causing engine knocking, as opposed to branched-chain octane isomers, which burn more slowly and give better performance. When used as a fuel component in antiknock test engines, a 100% heptane fuel is the zero point of the octane rating scale (the 100 point is 100% isooctane). Octane number equates to the antiknock qualities of a comparison mixture of heptane and isooctane, which is expressed as the percentage of isooctane in heptane and is listed in pumps for gasoline dispensed in the United States and internationally.

HF Alkylation: Alkylation using hydrofluoric acid as a catalyst.

High Pressure (HP): A processing unit operating at either equal to or greater than 225 psig measured at the outlet separator.

High Temperature Simulated Distillation (HTSD): Laboratory test designed for petroleum stocks boiling up to 1382°F (750°C).

HSR: Heavy Straight-Run. Usually a naphtha side stream from the atmospheric distillation tower.

HVGO: Heavy vacuum gas oil. A side stream from the vacuum distillation tower.

Hydrocarbon: Any organic compound that is comprised of hydrogen and carbon atoms, including crude oil, natural gas, and coal.

Hydrocrackate: The gasoline range product from a hydrocracker.

Hydrocracking: **1**. A process in which high or heavy gas oils or residue hydrocarbons are mixed with hydrogen under high pressure and temperature and in the presence of a catalyst to produce light oils. **2**. A refining process in which a heavy oil fraction or wax is treated with hydrogen over a catalyst under relatively high pressure and temperature to give products of lower molecular mass.

Hydrocracked Naphtha: A high-quality blending stream obtained when high-boiling cracked distillates undergo a combination of processes like cracking, hydrogenation, and reforming in the presence of a catalyst and hydrogen.

Hydrocyclone: A cone-shaped device for separating fluids and the solids dispersed in fluids.

Hydrodesulfurization: A process in which sulfur is removed from the molecules in a refinery stream by reacting it with hydrogen in the presence of a catalyst.

Hydrodesulfurizing: A process for combining hydrogen with sulfur in refinery petroleum streams to make hydrogen sulfide, which is removed from the oil as a gas.

Hydrogen: The lightest of all gases, the element (hydrogen) occurs chiefly in combination with oxygen in the water. It also exists in acids, bases, alcohols, petroleum, and other hydrocarbons.

Hydrogen Consumption: The amount of hydrogen that is consumed in a hydrocracking or hydrotreating process, usually expressed on a per-unit-of-feed basis. Hydrogen may be consumed in chemical reactions and dissolved and lost from the process in liquid hydrocarbon products.

Hydrogen Embrittlement: A corrosion mechanism in which atomic hydrogen enters between the grains of the steel and causes the steel to become very brittle.

Hydrogen-Induced Cracking: Stepwise internal cracks that connect hydrogen blisters.

Hydrogen Sulfide: **1**. "Rotten egg gas," H_2S. It is responsible for the distinctive odor of Rotorua. **2.** An objectionable impurity is present in some natural gas and crude oils and formed during the refining of sulfur-containing oils. It is removed from products by various treatment methods at the refining. **3**. Hydrogen sulfide is a gas with a rotten egg odor. This gas is produced under anaerobic conditions. Hydrogen sulfide gas is particularly dangerous because it dulls the sense of smell so that one does not notice it after one has been around it for a while. In high concentrations, hydrogen sulfide gas is only noticeable for a very short time before it dulls the sense of smell. The gas is very poisonous to the respiratory system, explosive, flammable, colorless, and heavier than air. **4**. A toxic, corrosive, colorless gas with the characteristic smell of rotten eggs in low concentration. An acid gas.

Hydrogen Sulfide Cracking: Minute cracking just under a metal's surface caused by exposure to hydrogen sulfide gas.

Hydrogenation: **1**. Filling in with hydrogen the "free" places around the double bonds in an unsaturated hydrocarbon molecule. **2**. A refinery process in which hydrogen is added to the molecules of unsaturated hydrocarbon fractions.

Hydrofining: A process of treating petroleum fractions and unfinished oils in the presence of catalysts and substantial quantities of hydrogen to upgrade their quality.

Hydroforming: A process in which naphtha is passed over a solid catalyst at elevated temperature and moderate pressures in the presence of added hydrogen to obtain high-octane motor fuels.

Hydroskimming Refinery: A topping refinery with a catalytic reformer.

Hydrostatic Pressure: Pressure created by a column of fluid that expresses uniform pressure in all directions at a specific depth and fluid composition above the measurement point.

Hydrotreating: **1**. A refinery process to remove sulfur and nitrogen from crude oil and other feedstocks. **2**. This is a term for a process by which product streams may be purified and otherwise be brought up to marketing specifications as to odor, color, stability, etc. **3**. A process in which a hydrocarbon is subjected to heat and pressure in the presence of a catalyst to remove sulfur and other contaminants such as nitrogen and metals and in which some hydrogenation can take place. Hydrotreating for the removal of sulfur is the major treating process in refineries. Cracked streams could be saturated and stabilized by converting olefins, albeit under more severe treating conditions. The process involves hydrogen under suitable temperature, pressure, and a catalyst.

Hyperforming: A catalytic hydrogenation process used for improving the octane number of naphtha by the removal of sulfur and nitrogen compounds.

H_2/Oil Ratio and Recycle Gas Rate: The H_2/oil ratio in standard cubic feet (scf) per barrel (bbl) is determined by:

$$\frac{H_2}{oil} = \frac{\text{total hydrogen gas to the reactor, scf / day}}{\text{total feed to the reactor, bbl / day}} [=] \frac{scf}{bbl}$$

An H_2/oil ratio in m^3/bbl is obtained by multiplying the H_2/oil ratio in (scf/bbl) by a conversion factor 0.028317. A molar H_2/oil ratio can be calculated from the volumetric H_2/oil ratio by the following equation:

$$\frac{\text{molar } H_2}{oil} = 1.78093 \times 10^{-7} \left(\frac{H_2}{oil} \; \frac{scf}{bbl} \right) \frac{MW_{oil}}{MW_{H_2}} \frac{\rho_{H_2}}{\rho_{oil}}$$

where MW_{oil} and MW_{H_2} are the molecular weights of the oil to be hydrotreated and of hydrogen, respectively, and ρ_{oil} and ρ_{H_2} are the densities of the oil and hydrogen (ρ_{H_2} at 15°C and 1 atm is 0.0898 kg/cm^2).

Hypothetical State: Is defined as a fluid in a state that cannot actually exist, e.g., methane as a liquid at 60°F and 14.696 psia. Methane cannot be in its liquid phase at this temperature and pressure, but such a state, when defined, can be used in calculations.

Identification and Structural Group Analysis: The crude oil is a complex mixture of saturated hydrocarbons; saturated hetero-compounds; and aromatic hydrocarbons, olefinic hydrocarbons, and aromatic hetero-compounds. With the advancement of instrumental analysis techniques like chromatography and spectroscopic methods, now, it is possible to study in depth the identification and structural group analysis. Some of the major analytical instruments used are gas chromatography, ion-exchange chromatography, simulated distillation by gas chromatography, absorption chromatography, gel permeation chromatography, high-performance liquid chromatography, and supercritical fluid chromatography. The application of spectroscopy, mass spectroscopy, electron spin resonance, X-ray diffraction, inductively coupled plasma emission spectroscopy, X-ray absorption spectroscopy, and atomic absorption spectrophotometer.

Initial boiling point (IBP): The initial boiling point of a cut, usually on an ASTM basis. The lowest temperature at which a petroleum product will begin to boil. The boiling temperature in a laboratory still at which the first drop of distilled liquid is condensed. The initial boiling point may be higher than the boiling point for light components in the sample that are not condensed by the apparatus.

Ignition: The process of starting a combustion process through the input of energy. Ignition occurs when the temperature of a substance is raised to the point at which its molecules will react spontaneously with an oxidizer and combustion occurs.

Ignition Quality: Ignition quality is very important in the case of high-speed automotive diesel engines. The diesel engine knock, engine noise, smoke, gaseous emissions, and so on, all depend upon this factor. Ignition quality is measured in terms of cetane number using an ASTM standard test engine. The test method designated as D613 comprises a single-cylinder engine with a variable compression ratio combustion pre-chamber.

Incident: See Accident.

Independent Protection Layer (IPL): Protection measures that reduce the level of risk of a serious event to 100 times, which have a high degree of availability (greater than 0.99) or have specificity, independence, dependability, and auditability.

Inerting: The process of removing an oxidizer (usually air or oxygen) to prevent a combustion process from occurring, normally accomplished by purging.

Inflammable: Has an identical meaning as flammable; however, the prefix "in" indicates a negative in many words and can cause confusion. Therefore, the use of flammable is preferred over inflammable.

Inherently Safer: **1**. A chemical process is inherently safer if it reduces or eliminates the hazards associated with the materials and operations used in the process, and this reduction or elimination is permanent and inseparable. **2**. An essential character of a process, system, or equipment that makes it without or very low in hazard or risk. Inherent safety is a way of looking at processes in order to achieve this. There are four main keywords:

- *Minimize (Intensification):* Reduce stocks of hazardous chemicals.
- *Substitute:* Replace hazardous chemicals with less hazardous ones.
- *Moderate (Attenuation):* Reduce the energy of the system—lowering pressures and temperatures or adding stabilizing additives generally make for lower hazards.
- *Simplify*: Make the plant and process simpler to design, build, and operate, hence making them less prone to equipment control and human failings.

Note: The principles of inherent safety are applied at the conceptual design stage to the proposed process chemistry. In certain instances, these hazards cannot be avoided; they are the basic properties of the materials and the conditions of usage. The inherently safer approach is to reduce the hazard by reducing the quantity of hazardous material or energy or by completely eliminating the hazardous agent.

Inherently Safer Design: Is a fundamentally different way of thinking about the design of chemical processes and plants. It focuses on the elimination or reduction of the hazards, rather than on management and control. This approach should result in safer and more robust processes, and it is likely that these inherently safer processes will also be more economical in the due course.

Instrument: The apparatus used in performing an action (typically found in instrumented systems).

Note: Instrumented systems in the process sector are typically composed of sensors (e.g., pressure, flow, temperature transmitters), logic solvers or control systems (e.g., programmable controllers, distributed control systems), and final elements (e.g., control valves). In special cases, instrumented systems can be safety instrumented systems.

Internal Combustion Engine (ICE): Is a heat engine where the combustion of a fuel occurs with an oxidizer (usually air) in a combustion chamber that is an integral part of the working fluid flow circuit. In an internal combustion engine, the expansion of the high-temperature and high-pressure gases produced by combustion apply direct force to some components of the engine. The force is applied typically to pistons, turbine blades, or a nozzle. This force moves the component over a distance, transforming chemical energy into useful mechanical energy (See Figure 20).

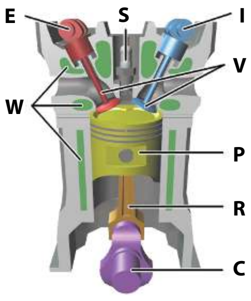


Figure 20 Diagram of a cylinder as found in four-stroke gasoline engines.

C	Crankshaft
E	Exhaust camshaft
I	Inlet camshaft
P	Piston
R	Connecting rod
S	Spark plug
V	Valves. Red: exhaust, blue: intake
W	Cooling water jacket
	Gray structure: Engine block.

Intrinsically Safe (IS): A circuit or device in which any spark or thermal effect is incapable of causing the ignition of a mixture of flammable or combustible materials in the air under prescribed test conditions.

IPTBE: Isopropyl tertiary butyl ether. An oxygenate used in motor fuels.

Isocracking: A hydrocracking process for the conversion of hydrocarbons to more valuable lower-boiling products by operation at relatively lower temperatures and pressures in the presence of hydrogen and a catalyst.

Isomerate: The product of an isomerization process.

Isomerization: 1. A refining process that alters the fundamental arrangement of atoms in the molecule without adding or removing anything from the original material. Used to convert normal butane into isobutane (iC_4H_{10}), an alkylation process feedstock, and normal pentane and hexane into isopentane (iC_5H_{12}) and isohexane (iC_6H_{14}), high-octane gasoline components. **2**. The rearrangement of straight-chain hydrocarbon molecules to form branched-chain products. Pentanes and hexanes, which are difficult to reform, are isomerized using precious metal catalysts to form the gasoline-blending components of a fairly high octane value. Normal butane may be isomerized to provide a portion of the isobutene feed needed for alkylation processes. The objective of isomerization is to convert low-octane n-paraffins to high-octane i-paraffins by using a chloride-promoted fixed bed reactor. **3**. Isomerization is the process by which one molecule is transformed into another molecule that has exactly the same atoms, but the atoms are rearranged. In some molecules and under some conditions, isomerization occurs spontaneously. Many isomers are equal or roughly equal in bond energy, and so they exist in roughly equal amounts, provided that they can interconvert relatively freely, that is, the energy barrier between the two isomers is not too high. When the isomerization occurs intermolecularly, it is considered a rearrangement reaction.

An example of an organometallic isomerization is the production of decaphenylferrocene [$(\eta^5\text{-}C_5\ Ph_5)_2Fe$] from its linkage isomer.

Isomers: Two compounds composed of identical atoms but with different structures/configurations, giving different physical properties. For example, hexane (C_6H_{14}) could be n-hexane, 2-methyl pentane, 3-methyl pentane, 2, 3-dimethyl butane, and 2, 2, - dimethylbutane.

A simple example of isomerism is given by propanol. It has the formula C_3H_8O (or C_3H_7OH) and occurs as two isomers: propanol-1-ol (n-propyl alcohol; II) and propanol-2-ol (isopropyl alcohol; III).

Note that the position of the oxygen atom differs between the two: It is attached to an end carbon in the first isomer and to the center carbon in the second (See Figures 21–23).

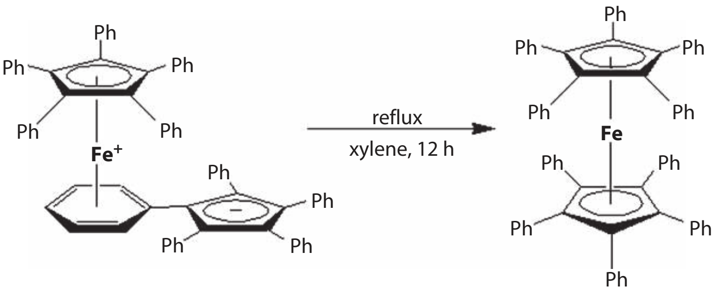


Figure 21

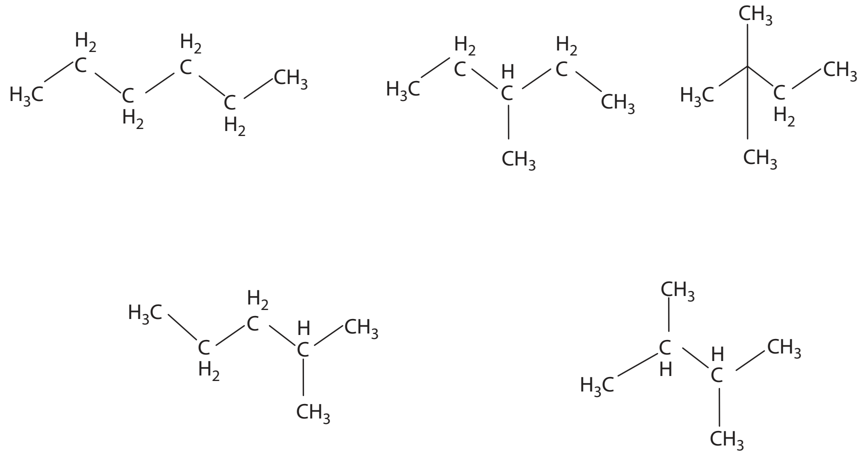


Figure 22

Isooctane – 2, 2, 4 – Trimethlypentane: Also known as isooctane or iso-octane, is an organic compound with the structure formula $(CH_3)_3CCH_2CH(CH_3)_2$.

It is one of several isomers of octane (C_8H_{18}). Engine knocking is an unwanted process that can occur during combustion in internal combustion engines.

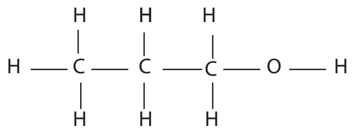
Graham Edgar, in 1926, added different amounts of n-heptane and 2,2,4 – trimethylpentane to gasoline and discovered that the knocking stopped when 2,2,4 trimethlypentane was added. Test motors, using 2,2,4 trimethylpentane, gave a certain performance that was standardized as 100 octane. The same test motors, run in the same fashion, using heptane, gave a performance that was standardized as 0 octane. All other compounds and blends of components then were graded against these two standards and assigned octane numbers. 2,2,4 trimethylpentane is the liquid used with normal heptanes (nC_7H_{18}*) to measure the octane number of gasoline. It is an important component of gasoline, frequently used in relatively large proportions to increase the knock resistance of the fuel (See Figure 24).*

Isopentane: See Natural Gasoline.

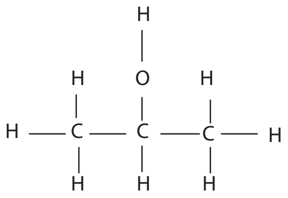
IVT: The initial vaporization temperature of a cut, usually based on a crude assay distillation.

Jack: An oil well pumping unit that operates with an up-and-down, or seesawing motion; also called a pumping jack.

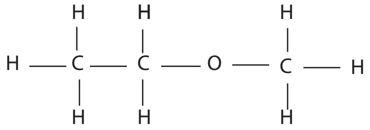
Jet fuel: A kerosene material of the typical ASTM D86 boiling point range 400–550°F (205–288°C) used as a fuel for commercial jet aircraft.



I



II



III

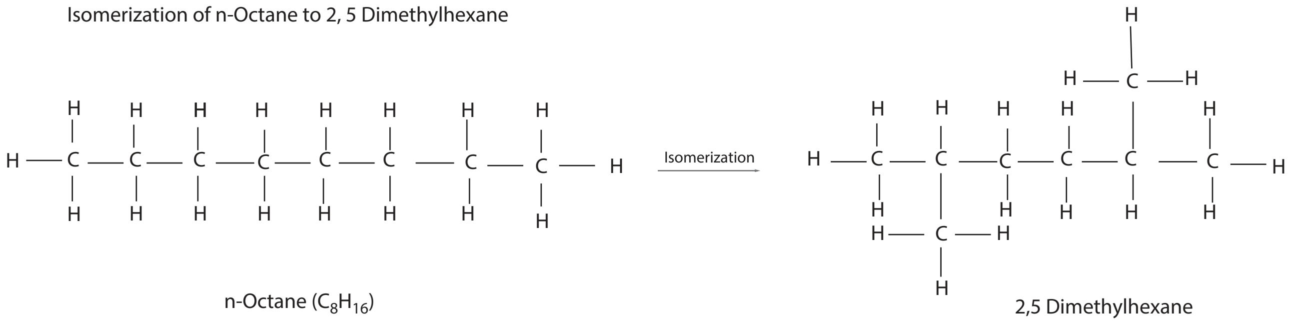


Figure 23

Joule–Thompson expansion: The pressure of a mixture is reduced with no heat transfer to or from the surroundings. A pressure decrease typically results in a temperature decrease, except for systems comprised largely of hydrogen gas.

Joule–Thompson Effect: **1**. The change in the temperature of a fluid that occurs when the fluid is allowed to expand in such a way that no external work is done and no heat transfer takes place. The case of most interest is the cooling of a compressed gas upon J-T expansion. NB: The J-T effect is not limited to gases; J-T expansion can, in some cases, produce an increase in temperature rather than a decrease, although this is not frequently encountered. **2.** The thermodynamic effect in a fluid whereby the reduction in its temperature is caused by pressure reduction without energy exchange with the environment. **3**. When a real (not ideal) gas expands, the temperature of the gas drops. During the passage of a gas through a choke, the internal energy is transferred to kinetic energy with a corresponding reduction in temperature as velocity increases. The effect for natural gas is approximately 7°F for every 100 psi pressure reduction.

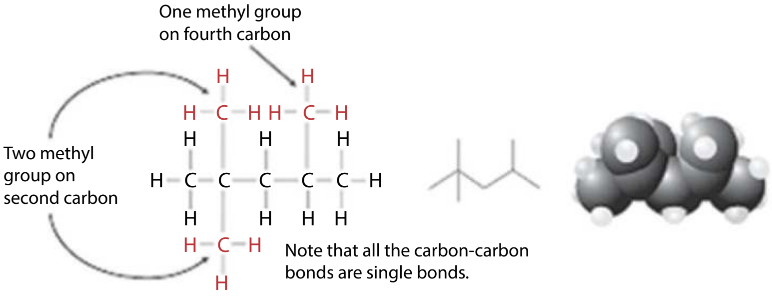


Figure 24

Joule–Thompson Valve: A device that, taking advantage of the Joule–Thompson effect, enables the cooling of fluid through throttling or the reduction of its pressure.

K factor: Sometimes used as a synonym for characterization factor.

K-value: Shortcut notation for the equilibrium K-value.

Kerogene: An initial stage of oil that never developed completely into crude. Typical of oil shales.

Kerosene/Kerosine: 1. A medium range (C_9–C_{16}) straight-chain blend of hydrocarbons. The flashpoint is about 140°F (60°C), the boiling point is 345–550°F (174–288°C), and the density is 747–775 kg/m^3. **2**. A medium–light distillate from the oil refining process; used for lighting and heating and for the manufacture of fuel for jet and turboprop aircraft engines. **3**. Any petroleum product with a boiling range between the approximate limits of 284 and 518°F (140 and 270°C), which satisfies specific quantity requirements. **4**. A middle distillate product material from the distillation of crude oil that boils in the approximate ASTM D86 range of 400–550°F (205–288°C) or from thermal and catalytic cracking operations (coker, visbreaker, FCC, hydrocracker, etc.). The exact cut is determined by various specifications of the finished kerosene. **5**. A light petroleum distillate that is used in space heaters, cook stoves, and water heaters and is suitable for use as a light source when burned in wick-fed lamps. Kerosene has a maximum distillation temperature of 400°F (204°C) at the 10% recovery point, a final boiling point of 572°F (300°C), and a minimum flash point of 100°F (38°C). Included are No. 1-K and No. 2-K, the two grades recognized by the American Society of Testing Materials (ASTM) Specification D3699 as well as all other grades of kerosene called range or stove oil, which have properties similar to those of No. 1 fuel oil. It is colorless and has a characteristic odor and taste. Kerosene is insoluble in water, moderately soluble in alcohol, and very soluble in ether, chloroform, or benzene.

Key Components: In a conventional distillation column with two products, two components or groups of components that define the separation. Both components must be distributed to the top and bottom products. The light key appears in the bottom product in a small significant quantity, and the heavy key appears in the top product in a small significant quantity.

Kinematic viscosity: Viscosity in centipoises (cP) divided by the liquid density at the same temperature gives kinematic viscosity in centistokes (cS) (100 cSt = 1 stoke). Water is the primary viscosity standard with an accepted viscosity at 20°C of 0.01002 poise. Kinematic viscosity is usually determined by the flow of a substance between two points in a capillary tube.

$$\text{Kinematic viscosity} = \frac{\text{Dynamic viscosity}}{\text{Density of fluid}}, \text{cSt}$$

$$\nu = \frac{\mu}{\rho}$$

Kinetic: The word “kinetic” is derived from the Greek word for “motion.” In chemistry, kinetics is the study of how fast reactions occur. In many chemical reactions where there are a number of possible products, the first product formed may be the one that is formed most quickly, not necessarily the one that is most stable; if the reaction is left to proceed, eventually, a product is formed that involves the greatest change in bond energy—the thermodynamic product.

Knock: **1**. The sound associated with the autoignition in the combustion chamber of an automobile engine of a portion of the fuel–air mixture ahead of the advancing flame front. **2**. The noise is associated with premature ignition of the fuel–air mixture in the combustion chamber; also known as detonation or pinking.

Knocking (knock, detonation, spark knock, pinging, or pinking): In spark-ignition internal combustion engines, knocking occurs when the combustion of the air/fuel mixture in the cylinder does not start off correctly in response to ignition by the spark plug, but one or more pockets of air/fuel mixture explode outside the envelope of the normal combustion front. See also Engine Knocking.

Knocking is more or less unavoidable in diesel engines, where fuel is injected into highly compressed air toward the end of the compression stroke. There is a short lag between the fuel being injected and combustion starting. By this time, there is already a quantity of fuel in the combustion chamber that will ignite first in areas of greater oxygen density prior to the combustion of the complete charge. This sudden increase in pressure and temperature causes the distinctive diesel "knock" or "clatter," some of which must be allowed for in the engine design. A careful design of the injector pump, fuel injector, combustion chamber, piston crown, and cylinder head can reduce knocking greatly, and modern engines using electronic common rail injection have very low levels of knock. Engines using indirect injection generally have lower levels of knock than direct injection engines due to the greater dispersal of oxygen in the combustion chamber and lower injection pressures, providing a more complete mixing of fuel and air.

Knocking should not be confused with pre-ignition—they are two separate events. However, pre-ignition is usually followed by knocking. See Pre-ignition.

Knockout: A separator used to remove easily removed or excess gas or water from the produced fluid stream.

Knockout Drum: A vessel wherein suspended liquid is separated from gas or vapor.

Laminar flow: The streamlined flow of a fluid in which a fluid flows without fluctuations or turbulence. The velocities of fluid molecules are in the direction of flow with only minor movement across the streamlines caused by molecular diffusion. The existence was first demonstrated by Osborne Reynolds who injected a trace of colored fluid into a flow of water in a glass pipe. At low flow rates, the colored fluid was observed to remain as discrete filaments along the tube axis, indicating flow in parallel streams. At increased flow rates, oscillations were observed in the filaments, which eventually broke up and dispersed across the tube. There appeared to be a critical point for a particular tube and fluid above which the oscillations occurred. By varying the various parameters, Reynolds showed that the results could be correlated into the terms of a dimensionless number called the Reynolds number, Re. This is expressed by:

$$\mathrm{Re} = \frac{\rho v d}{\mu}$$

where ρ is the density of the fluid, v is the velocity of the fluid, d is the inside diameter of the pipe, and μ is the fluid viscosity. The critical value of Re for the break-up of laminar flow in the pipes of circular cross-section is about 2000.

Leaded Gasoline: A gasoline that has TEL (tetraethyl lead) added to boost the octane number.

LCGO: Light coker gas oil

Lean Oil: The absorption oil entering the top tray of an absorber column.

Lease Condensate: A mixture consisting primarily of pentanes and heavier hydrocarbons that are recovered as a liquid from natural gas in lease separation facilities. This category excludes natural gas liquids, such as propane and butane, which are recovered at downstream natural gas processing plants or facilities. *See also Natural Gas Liquids.*

LHSV: Liquid hour space velocity, the volume of feed per hour per volume of a catalyst.

LHV: The lower heating value of fuels (net heat of combustion). The heat produced by complete oxidation of materials at 60°F (25°C) to carbon dioxide and water vapor at 60°F (25°C).

Light Cycle Oil (LCO): Gas oil produced in a catalytic cracking operation that boils in the approximate ASTM D86 range of 400–695°F (205–369°C).

Light Ends: Hydrocarbon fractions in the butane (C_4H_{10}) and lighter boiling range.

Light Gas Oils: Liquid petroleum distillates heavier than naphtha, with an approximate boiling range from 401 to 650°F (205–343°C).

Light key: A distributed component in a distillation section that is recovered in the top light product, with a small specified amount leaving the bottom product.

Light oil: Generally gasoline, kerosene, and distillate fuels.

Light Straight-Run (LSR): The low-boiling naphtha stream from the atmospheric distillation, usually composed of pentanes and hexanes.

Liquefaction: **1**. The process by which gaseous natural gas is converted into liquid natural gas. **2**. The physical process from gas to liquid is condensation. For natural gas, this process requires cryogenic temperature since it is impossible to liquefy methane—the main component of natural gas—at a temperature above -117°F (-82.6°C), which is its critical temperature.

Liquefaction of Gases: Any process in which a gas is converted from a gaseous-to-liquid phase.

Liquefaction Plant: Industrial complex that processes natural gas into LNG by removing contaminants and cooling the natural gas into its condensation.

Liquefaction Unit or Liquefaction Train: Equipment that processes purified natural gas and brings it to a liquid state. Natural gas has been purified in the pretreatment unit before cooling and liquefying it.

Liquefied Natural Gas (LNG): **1**. Natural gas that has been refrigerated to temperatures at which it exists in a liquid state. **2**. An odorless, colorless, non-corrosive, and non-toxic product of natural gas consisting primarily of methane (CH_4) that is in liquid form at near-atmospheric pressure. **3**. Natural gas liquefied either by refrigeration or by pressure to facilitate storage or transportation. **4**. A liquid composed chiefly of natural gas (e.g., mostly methane, CH_4). Natural gas is liquefied to make it easy to transport if a pipeline is not feasible (e.g., as across a body of water). LNG must be put under low temperature and high pressure or under extremely low (cryogenic) temperature and close to atmospheric pressure to liquefy. **5**. Natural gas, mainly methane, refrigerated to reach a liquid phase suitable for transportation in specialized vessels. **6**. Natural gas that has been cooled to -26°F (-32°C) and converted into a liquid so that its volume will be reduced for transportation. **7**. A hydrocarbon mixture, predominantly methane, kept in a liquid state at a temperature below its boiling point. **8**. Methane that has been compressed and cooled to the liquefaction point for shipping.

Liquefied Petroleum Gas (LPG): **1**. Gaseous hydrocarbons at normal temperatures and pressures but that readily turns into liquids under moderate pressure at normal temperatures, i.e., propane, (C_3H_8) and butane (C_4H_{10}). **2**. Butane and propane mixture, separated from well fluid stream. LPG can be transported under pressure in refrigerated vessels (LPG carriers). **3**. A mixture of propane and butane and other light hydrocarbons derived from refining crude oil. At normal temperatures, it is a gas, but it can be cooled or subjected to pressure to facilitate storage and transportation. **4**. Of the gaseous hydrocarbons, propane and butane can be liquefied under relatively low pressure and at ambient temperature. Mixtures of these are known as LPG. **5**. A mixture of propane, propylene, butane, and butylenes. When compressed moderately at normal temperature, it becomes a liquid. It is obtained as light ends from the fractionation of crude oil. It has a good caloric value; it is used as cooking fuel. Because LPG has no natural odor, a distinctive odorant is added so that it will be noticeable should a leak occur. **6**. Light ends, usually C_3 and C_4 gases liquefied for storage and transport. **7**. Propane, propylene, normal butane, butylenes, isobutane, and isobutylene produced at refineries or natural gas processing plants (includes plants that fractionate raw natural gas plant liquids). **8**. A group of hydrocarbon-based gases derived from crude oil refining or natural gas fractionation. They include ethane, ethylene, propane, propylene, normal butane, butylenes, isobutane, and isobutylene. For the convenience of transportation, these gases are liquefied through pressurization.

Liquefied Refinery Gases: Liquefied petroleum gases fractionated from refinery or still gases. Through compression and/or refrigeration, they are retained in a liquid state. The reported categories are ethane/ethylene, propane/propylene, normal butane/butylenes, and isobutane/isobutylene.

Liquid Extraction: Light and heavy liquid phases are contacted in a column with contact surfaces and possibly mixing. Some components are transferred (extracted) from one liquid phase to the other.

Lockout–Tagout (LOTO): Refers to a program to control hazardous energy during the servicing and maintenance of machinery and equipment. Lockout refers to the placement of a locking mechanism on an energy-isolating device, such as a valve, so that the equipment cannot be operated until the mechanism is removed. Tagout refers to the secure placement of a tag on the energy-isolating device to indicate that the equipment cannot be operated until the tag is removed (See Figure 25).

Long Residue: The bottoms stream from the atmospheric distillation tower.

Long-Term Exposure Limit (LTEL): The time-weighted average concentration of a substance over an 8-h period thought not to be injurious to health.

Lower Explosive Limit (LEL): The minimum concentration of combustible gas or vapor in air below which the propagation of flame does not occur on contact with an ignition source. Also known as the lower flammable limit or the lower explosion limit.

Low Pressure (LP): A processing unit operating at less than 225 psig measured at the outlet separator.

Lubricants: Substances used to reduce the friction between bearing surfaces or as process materials either incorporated into other materials used as processing aids in the manufacture of other products or used as the carriers of other materials. Petroleum lubricants may be produced either from distillates or residues. Lubricants include all grades of lubricating oils, from spindle oil to cylinder oil and those used in greases.

Light Vacuum Gas Oil (LVGO): A side stream from the vacuum distillation tower.

Figure 25 An isolation tag on a piece of equipment (Source: hazardex, www.hazardexonthenet.net).

Make Up Stream: A feed to a process to replace a component that reacts or is otherwise depleted in a process.

Main Cryogenic Heat Exchanger: Main heat exchanger in the liquefaction unit where the cooling and liquefaction of natural gas take place by means of heat exchange with cooling fluids.

Main Fractionators: The first distillation column for an FCC or coking process.

Main Line: Branch or lateral sewers that collect wastewater from building sewers and service lines.

Main Sewers: A sewer that receives wastewater from many tributary branches and sewer lines and serves as an outlet for a large territory or is used to feed an intercepting sewer.

Management of Change (MOC): 1. A process to understand all the implications of a change to a procedure. **2.** A process for evaluating and controlling hazards that may be introduced during modifications to the facility, equipment, operations, personnel, or activities; MOCs can also be used to identify, evaluate, and control unintended hazards introduced by modifying procedures or when developing a new plan or procedure.

Manhole: An opening in a sewer provided for the purpose of permitting operators or equipment to enter or leave a sewer. Sometimes called an "access hole" or a "maintenance hole."

Manifold(s): 1. A junction or center for connecting several pipes and selectively routing the flow. **2**. A pipe spool in which a number of incoming pipes are combined to feed to a common output line.

Manometer: Instrument for measuring head or pressure; basically, a U-tube partially filled with a liquid, so constructed that the difference in the level of the liquid leg indicates the pressure exerted on the instrument.

MAOP: See Maximum Allowable Operating Pressure.

Mass Balance: The summation of the mass entering a process and the summation of the mass leaving a process. They must be equal for a steady-state process.

Material Safety Data Sheet: **1**. A description of the Health, Safety, and Environment (HSE) data for a marketed product. **2**. Printed information that describes the properties of a hazardous chemical and ways to control its hazards. **3**. A document that provides pertinent information and a profile of a particular hazardous substance or mixture. An MSDS is normally developed by the manufacturer or formulator of the hazardous substance or mixture. The MSDS is required to be made available to employees and operators whenever there is the likelihood of the hazardous substance or mixture being introduced into the workplace.

MAWP: *See Maximum Allowable Working Pressure*.

Maximum Allowable Operating Pressure (MAOP): The maximum gas pressure at which a pipeline system or process facility is allowed to operate.

Maximum Allowable Working Pressure (MAWP): **1**. This is a legal maximum pressure that a process vessel is allowed to experience. Above this pressure, a relief valve should be open to protect the vessel from catastrophic failure. **2**. The maximum pressure to which a surface vessel can be operated or the maximum pressure during treating to which a well should be exposed.

Mechanical Seal: This is the part of a centrifugal pump that keeps the liquid from squirting out along the shaft. It is often subject to leakage due to pump vibration and cavitation.

Melting Point: The temperature at which a solid turns into a liquid. As the temperature is a measure of the kinetic energy of molecules (i.e., how much they are moving around), this means that the molecules are moving too much to stay in one place.

Mercaptans: **1**. Compounds of carbon, hydrogen, and sulfur (RSH, R=CH_3) found in sour crude and gas; the lower mercaptans have a strong, repulsive odor and are used, among other things, to odorize natural gas. **2**. A class of compounds containing carbon, hydrogen, and sulfur. The shorter-chain materials are used as an odor marker in natural gas. **3**. Organic sulfides of the formula RSH where R represents the organic radical and SH represents the thiol group.

Methane (CH_4): A light odorless flammable gas that is the principal component of natural gas.

Methanol (CH_3OH): Methyl alcohol from the general formula (ROH), where $R = C_nH_{2n+1}$ is known as a radical and n = 1, 2, 3, etc. Methanol can be made by the destructive distillation of wood or through a process starting with methane or a heavier hydrocarbon, decomposing it to synthesis gas and recombining it to methanol.

Methyl Tertiary Butyl Ether (MTBE, - $(CH_3)_3COCH_3$): **1**. Is manufactured by the etherification of methanol and isobutylene. Methanol is derived from natural gas, and isobutylene is derived from butane obtained from crude oil and natural gas **2**. A gasoline additive used to increase the octane number. MTBE is produced by reacting methanol (CH_3OH) with isobutylene (iC_4H_8). **3**. Blends up to 15.0% by volume of MTBE that must meet the ASTM D4814 specifications. Blenders must take precautions that the blends are not used as base gasoline for other oxygenated blends (commonly referred to as the "Sun waiver").

An ether intended for gasoline blending as described in oxygenate definition.

In the United States, it has been used in gasoline at low levels since 1979 to replace tetraethyl lead and to increase its octane rating, helping to prevent engine knocking. Oxygenates help gasoline burn more completely, reducing tailpipe emissions from pre-1984 motor vehicles; dilutes or displaces gasoline components such as aromatics (e.g., benzene) and sulfur; and optimizes the oxidation during combustion. Most refiners chose MTBE over other oxygenates primarily for its blending characteristics and low cost.

Middle Distillates: Atmospheric pipe still cuts boiling in the range of 300–700°F (149–371°C) vaporization temperature. The exact cut is determined by the specifications of the product. **1**. A general classification of refined petroleum products that include distillate fuel oil and kerosene. **2**. Medium-density refined petroleum products, including kerosene, stove oil, jet fuel, and light fuel oil. **3**. Refinery products in the middle distillation range of refined products: kerosene, heating oil, and jet fuel.

Mid-Percent Point: The vapor temperature at which one-half of the material of a cut has been vaporized. Mid-percent point is used to characterize a cut in place of temperature limits.

Mixed Phase: More than one phase. Usually implies both vapor and liquid phase(s) present.

Molecular Sieve: A separation process that usually works by gaseous diffusion. A membrane is selected through which the compounds being removed or purified can pass, while the remaining compounds in the stream being processed cannot pass.

MONC: Motor octane number clear (unleaded).

Motor Octane Number (MON, ASTM ON F2): A measure of resistance to the self-ignition (knocking) of gasoline under laboratory conditions that correlate with road performance during highway driving conditions. The percentage by volume of isooctane in a mixture of isooctane and n-heptane that knocks with the same intensity as the fuel being tested. A standardized test engine operating under standardized conditions (900 rpm) is used. This test approximates the cruising conditions of an automobile; ASTM D-2723.

MPHC: Medium-pressure hydrocracking or partial conversion hydrocracking.

Motor Gasoline or Petrol: Gasoline is a volatile, flammable, complex petroleum fuel used mainly in internal combustion engines. It is used as fuel in specially designed heaters and lamps.

Motor Gasoline Blending: **1**. Naphthas (e.g., straight-run gasoline, alkylate, reformate, benzene, toluene, xylenes) used for blending or compounding into finished motor gasoline. Includes receipts and inputs of Gasoline Treated as Blendstock (GTAB). Excludes conventional blendstock for oxygenate blending (CBOB), reformulated blendstock for oxygenate blending, oxygenates (e.g., fuel ethanol and methyl tertiary butyl ether), butane, and natural gasoline. **2**. Mechanical mixing of motor gasoline blending components, and oxygenates when required, to produce finished motor gasoline. Finished motor gasoline may be further mixed with other motor gasoline blending components or oxygenates, resulting in increased volumes of finished motor gasoline and/or changes in the formulation of finished motor gasoline (e.g., conventional motor gasoline mixed with MTBE to produce oxygenated motor gasoline).

Motor Gasoline Blending Components: Naphthas (e.g., straight-run gasoline, alkylate, reformate, benzene, toluene, xylene) used for blending or compounding into finished motor gasoline. These components include reformulated gasoline blendstock for oxygenate blending (RBOB) but exclude oxygenates (alcohols, ethers), butane, and pentanes plus.

Motor gasoline (finished): A complex mixture of relatively volatile hydrocarbons with or without small quantities of additives, blended to form a fuel suitable for use in spark-ignition engines. Motor gasoline, as defined in ASTM Specification D4814 or Federal Specification VV – G – 1690C, is characterized as having a boiling range of 122–158°F (50–70°C) at the 10% recovery point to 365–374°F (185–190°C) at the 90% recovery point. "Motor gasoline" includes conventional gasoline, all types of oxygenated gasoline, including gasohol, and reformulated gasoline but excludes aviation gasoline.

Naphtha: **1**. Straight-run gasoline distillate, below the boiling point of kerosene. Naphthas are generally unsuitable for blending as a component of premium gasoline; hence, they are used as a feedstock for catalytic reforming in hydrocarbon production processes or in chemical manufacturing processes. **2**. A term that is applied to low-boiling mixtures of hydrocarbons with typical TBP boiling ranges between 150 and 450°F (66–233°C). Light and heavy naphthas are produced in the distillation of crude oils. Cracked naphthas are also produced by many of the refinery reaction processes.

Naphthas are subdivided according to the actual pipe still cuts—into light, intermediate and heavy, and very heavy virgin naphthas. A typical pipe still operation would be

C_5–160°F (C_5–71°C) : light virgin naphtha
160–280°F (71–138°C) : intermediate virgin naphtha
280–380°F (138–193°C) : heavy virgin naphtha

Naphtha, the major constituent of gasoline, generally needs processing to make suitable quality gasoline.

Naphtha Less Than 401°F: A naphtha with a boiling range of less than 401°F (205°C) that is intended for use as a petrochemical feedstock.

Naphtha-Type Jet Fuel: A fuel in the heavy naphtha boiling range having an average gravity of 52.8 °API, 20–90% distillation temperature of 290–470°F (143–243°C), and meeting Military Specification MIL – T- 5624L (Grade JP–4). It is used primarily for military turbojet and turboprop aircraft engines because it has a lower freeze point than other aviation fuels and meets engine requirements at high altitudes and speeds.

Special Naphthas: All finished products within the naphtha boiling range that are used as paint thinners, cleaners, or solvents. These products are refined to a specified flash point. Special naphthas include all commercial hexane and cleaning solvents conforming to ASTM Specification D 1836 and D484, respectively. Naphthas to be blended or marketed as motor gasoline or aviation gasoline and synthetic natural gas (SNG) feedstocks are excluded.

Naphthenes: Hydrocarbons of the cyclane family, sometimes called cycloalkanes. Naphthenes have no double bonds and are saturated ring structures with the general formula C_nH_{2n}, where C = carbon atoms, H = hydrogen atoms, and n = 6, 7, 8, …

Naphthenic: Having the characteristics of naphthenes, saturated hydrocarbons whose molecules contain at least one closed ring of carbon atoms.

Naphthenic Acids: Organic acids occurring in petroleum that contain a naphthenic ring and one or more carboxylic acid groups. Naphthenic acids are used in the manufacture of paint driers and industrial soaps.

Naphthenic Crudes: A type of crude petroleum containing a relatively large proportion of naphthenic-type hydrocarbon.

Natural Gas: Naturally occurring gas consisting predominantly of methane, sometimes in conjunction with crude (associated gas) and sometimes alone (unassociated gas). **1**. A mixture of light hydrocarbons found naturally in the Earth's crust, often in association with oil (when it is known as associated gas). Methane is the most dominant component. It may also include some short-chain hydrocarbons (ethane, propane, butane) that may be in a gaseous state at standard conditions. **2**. A mixture of hydrocarbon compounds and small quantities of various non-hydrocarbons existing in the gaseous phase or in a solution with crude oil in natural underground reservoirs at reservoir conditions. The primary constituent compound is CH_4. Gas coming from wells also can contain significant amounts of ethane (C_2H_6), propane (C_3H_8), butane (C_4H_{10}), and pentanes (C_5H_{12}) and widely varying amounts of carbon dioxide (CO_2) and nitrogen (N_2).

Natural Gas Heating Value: The amount of thermal energy released by the complete combustion of one standard cubic foot of natural gas.

Natural Gas Liquids (NGL): 1. Liquid hydrocarbons, such as ethane, propane, butane, pentane, and natural gasoline, extracted from field natural gas. **2**. Those hydrocarbons in natural gas that are separated from the gas as liquids through the process of absorption, condensation, adsorption, or other methods of gas processing or cycling plants. Generally, such liquids consist of propane and heavier hydrocarbons and are commonly referred to as lease condensate, natural gasoline, and liquefied petroleum gases. Natural gas liquids include natural gas plant liquids (primarily ethane, propane, butane, and isobutane). See Natural Gas Plant Liquids and Lease Condensate (primarily pentanes produced from natural gas and lease separators and field facilities). **3**. Liquids obtained during natural gas production include ethane, propane, butanes, and condensate.

Natural Gasoline: A gasoline range product separated at a location near the point of production from natural gas streams and used as a gasoline blending component.

Natural Gasoline and Isopentane: A mixture of hydrocarbons, mostly pentanes and heavier, extracted from natural gas, that meets vapor pressure, and endpoint and other specifications for natural gasoline set by the Gas Processors Association. Includes isopentane that is a saturated branch-chain hydrocarbon (iC_5H_{12}), obtained by the fractionation of natural gasoline or the isomerization of normal pentane (nC_5H_{12}).

Natural Gas Plant Liquids: Those hydrocarbons in natural gas that are separated as liquids at natural gas processing plants, fractionating and cycling plants, and, in some instances, field facilities. Lease condensate is excluded. Products obtained include ethane, liquefied petroleum gases (propane, butanes, propane–butane mixture, ethane–propane mixture), isopentane, and other small quantities of finished products, such as motor gasoline, special naphthas, jet fuel, kerosene, and distillate fuel oil.

Natural Gas Processing: **1**. The purification of field gas at natural gas processing plants (or gas plants) or the fractionation of mixed NGLs to natural gas products to meet specifications for use of pipeline-quality gas. Gas processing includes removing liquids, solids, and vapors absorbing impurities and odorizing. **2**. The process of separating natural gas liquids (NGLs) by absorption, adsorption, refrigeration, or cryogenics from the steam of natural gas.

Natural Gas Processing Plant: Facilities designed to recover natural gas liquids from a stream of natural gas that may or may not have passed through lease separators and/or field separation facilities. These facilities control the quality of the natural gas to be marketed. Cycling plants are classified as gas processing plants.

Net Heating Value: Is the total energy transferred as heat in an ideal combustion reaction at a standard temperature and pressure in which all water formed appears as vapor. The net heating is an ideal gas property in a hypothetical state (the water cannot all remain vapor because after the water saturates the CO_2 in the products, the rest would condense).

Net Positive Suction Head (NPSH): The net-positive suction head required to keep a centrifugal pump from cavitating. Cooling a liquid in a pump's suction line increases the pump's available NPSH, as does increasing the liquid level in the suction drum.

Net Zero: Refers to a point at which the amount of greenhouse gas being put into the atmosphere by human activity equals the amount of greenhouse gas that is being taken out of the atmosphere.

Non-associated Gas: Natural gas that exists in a reservoir alone and is produced without any crude oil.

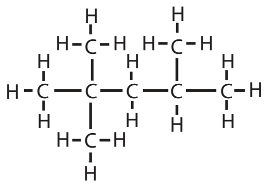
Normal Boiling Point: *See Boiling Point*.

Nusselt Number (Nu): A dimensionless number Nu is used in heat transfer calculations characterizing the relation between the convective heat transfer of the boundary layer of a fluid and its thermal conductivity:

$$\mathrm{Nu} = \frac{hd}{k}$$

where h is the surface heat transfer coefficient, d is the thickness of the fluid film, and k is the thermal conductivity.

Octane (C_8H_{18}): 1. Is a hydrocarbon and an alkane with the chemical formula C_8H_{18}, and the condensed structural formula $CH_3(CH_2)_6CH_3$. Octane has many structural isomers that differ by the amount and location of branching in the carbon chain. One of the isomers, 2, 2, 4 – trimethylpentane (isooctane) $(CH_3)_3CCH_2CH(CH_3)_2$



is used as one of the standard values in the octane rating scale. Octane is a component of gasoline (petrol). As with all low-molecular-weight hydrocarbons, octane is volatile and very flammable. **2**. A test used to measure the suitability of gasoline as motor fuel. The octane test determines the knocking characteristics of gasoline in a standard test engine relative to a standard of 2-2-4 trimethyl pentane (2 2 4 TMP). 2 2 4 TMP is assigned an octane number of 100.0. There are two octane tests. One is designated the research octane (F-1), and the second is the motor octane (F-2). Motor octane is determined in an engine more representative of actual operating conditions for automobiles and is lower than research octane for any gasoline stock.

Historically, gasoline was marketed based on the F-1 octane, but in recent years, the average of the F-1 and F-2 octane has been used. At the gasoline pump, this is reported as (R + M)/2.

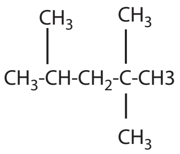
The Research Octane Number (RON) test simulates driving mild conditions, while the Motor Octane Number (MON) test simulates driving under more severe conditions, i.e., under load and at high speeds. The arithmetic average of RON and MON, which gives an indication of the performance of the engine under the full range of conditions, is projected as Antiknock Index (AKI), i.e., Antiknock Index (AKI) = (RON + MON)/2.

Octane Number: **1**. Is a measure of the knocking characteristics of fuel in a laboratory gasoline engine according to ASTM D2700. We determine the octane number of fuel by measuring its knocking value compared to the knocking of a mixture of n-heptane and isooctane of 2-2-4 trimethylpentane (224 TMP). **2**. An index measured by finding a blend of isooctane (iC_8H_{18}) and normal heptanes (nC_7H_{16}) that knocks under identical conditions as the gasoline being evaluated. It is a measure of the ease of self-ignition of fuel without the aid of a spark plug. **3**. The octane number is a measure of the antiknock resistance of gasoline. It is the percentage of isooctane in a mixture of isooctane and n-heptane, which gives a knock of the same intensity as the fuel being measured when compared in a standard engine. For example, if the fuel being tested matches in knocking to a blend of 90% isooctane and 10% n-heptane, then the test fuel is said to have an octane number of 90.

Iso-octane, which produces the least knocking or which knocks only at a much higher compression ratio, is given an octane number of 100, while n-heptane, which is very poor in its resistance to knocking or which knocks at a much lower compression ratio, is given an octane number of zero.

$CH_3CH_2\text{-}CH_2\text{-}CH_2\text{-}CH_2\text{-}CH_2\text{-}CH_3$ [Octane No. = 0]

n-heptane



$$\overset{CH_3}{\overset{|}{CH_3\text{-}CH}}\text{-}CH_2\text{-}\underset{\underset{CH_3}{|}}{\overset{\overset{CH_3}{|}}{C}}\text{-}CH3$$

[Octane No. = 100]

Iso-octane
[2,2,4 -trimethyl pentane]

Generally, the octane number increases as the degree of branching of the carbon chain increases, and thus isoparaffins are found to give higher octane numbers than the corresponding normal isomers. Olefins are found to give higher octane numbers than the related paraffins. Naphthenes also give better octane numbers than the corresponding normal paraffins. Aromatics usually exhibit high octane numbers.

A single-cylinder test engine is made to obtain the antiknock characteristics of gasoline in terms of octane numbers. The octane numbers formed a scale ranging from 0 to 100; the higher the number, the greater the antiknock characteristics. The scale has been extended above 100 by comparing the knocking intensity with isooctane to which tetraethyl lead (TEL) is added. Numbers greater than 100 on the scale are referred to as performance numbers rather than octane numbers. *See also Motor Octane Number and Research Octane Number (Table 2).*

Octane numbers are very relevant in the reforming, isomerization, and alkylation processes in refining facilities. These processes enable the successful reactive transformations to yield long side-chain paraffin and aromatics that possess higher octane numbers than the feed constituents that do not consist of higher quantities of constituents possessing straight-chain paraffin and non-aromatics (naphthenes).

It is a measure of the ease of self-ignition of fuel without the aid of a spark plug.

Octane Scale: A series of arbitrary numbers from 0 to 120.3 used to rate the octane number of gasoline. Three reference materials define the scale; n-heptane (Octane number = 0), isooctane (Octane number = 100), and isooctane plus 6 ml tetramethyl lead (Octane number = 120.3). Above 100, the octane number of a fuel is based on the engine ratings, in terms of the ml of tetra ethyl lead in isooctane, which matches that of the unknown fuel.

Off Gas: The gas leaving a reflux drum or top tray of an absorber column.

Offline: When a process unit is shut down, it is said to be offline.

Oil: One of the various liquid, viscid, usually inflammable, chemically neutral substances that is lighter and insoluble in water but soluble in alcohol and ether and classified as non-volatile. Natural plant oils comprise terpenes and simple esters such as essential oils. Animal oils are glycerides of fatty acids. Mineral oils are mixtures of hydrocarbons. Oils have many uses and include fuels lubricants, soap constituents, vanishes, etc.

Oil and Gas: Refer to the industry associated with the recovery of liquid and gaseous hydrocarbons from underground deposits as reservoirs found both onshore and offshore around the world. A collection of localized deposits is known as an oil field or gas field. When they are drilled, they are known as oil and gas wells. Oil is mainly used as fuel for transportation purposes, whereas gas is primarily used as fuel for domestic and industrial purposes and for converting into other chemicals such as plastic. Oil is widely transported in ships. Gas is transported in underground, sub-sea, or overland pipelines covering large distances.

Table 2 Octane numbers of pure hydrocarbons*.

Hydrocarbon	RON	MON	Hydrocarbon	RON	MON
n-Pentane	61.7	61.9	2,4 – Dimethyl hexane	62.5	69.9
n-Hexane	24.8	26.0	2,2,4 –Trimethyl pentane (isooctane)	100.0	100.0
n-Heptane	0.0	0.0	1 – Pentene	90.9	77.1
n-Octane	-19.5	-15.0	1 – Octane	28.7	34.7
n-Nonane	-17.0	-20.0	3 - Octene	72.5	68.1
2-Methyl butane (iso-pentane)	92.3	90.3	4 – Methyl – 1- Pentene	95.7	80.9
2 – Methyl hexane (iso-heptane)	42.4	46.4	Benzene	-	114.8
2 – Methyl heptane (iso-octane)	21.7	23.8	Toluene	120.1	103.5

*(Source: Speight, James G., The Chemistry & Technology of Petroleum, Marcel Dekker, Inc. 1991).

Oil Refinery: An industrial process plant where crude oil is converted into useful products such as naphtha, diesel fuel, kerosene, and LPG. Also known as petroleum refinery, the process involves the separation of the crude oil into fractions in the process of fractional distillation. By boiling the crude oil, the light or more volatile components with the lowest boiling point rise toward the top of the column, whereas the heavy fractions with the highest boiling points remain at the bottom. The heavy bottom fractions are then thermally cracked to form more useful light products. All the fractions are then processed further in other parts of the oil refinery, which may typically feature vacuum distillation used to distill the bottoms; hydrotreating, which is used to remove sulfur from naphtha, catalytic cracking, fluid catalytic cracking, hydrocracking, visbreaking, isomerization, steam reforming, alkylation, hydrodesulfurization, and the Claus process used to convert hydrogen sulfide into sulfur, solvent dewaxing, and water treatment (See Figure 26).

Olefins: Hydrocarbons of the alkenes family. Olefins have two carbon atoms in the molecular structure linked by a double bond to satisfy the absence of two hydrogen atoms that are present in the corresponding paraffin. This hydrogen deficiency is called unsaturation. The general formula for olefins is C_nH_{2n}, where C = carbon atoms, H = hydrogen atoms, and n = 2, 4, 6…

Olefins do not occur naturally in crude oil and are created in the thermal and catalytic cracking processes.

Online: When a process unit is in operation and the processing feed, it is said to be online.

OPEC: Organization of Petroleum Exporting Countries. These countries have organized for the purpose of negotiating with oil companies on matters of oil production, prices, and future concession rights. The current members are Algeria, Indonesia, Iran, Kuwait, Libya, Nigeria, Qatar, Saudi Arabia, United Arab Emirates, and Venezuela.

Operability Capacity: **1**. The amount of capacity that, at the beginning of the period, is in operation; not in operation and not under active repair but capable of being placed in operation within 30 days; or not in operation but under active repair that can be completed within 90 days. Operable capacity is the sum of the operating and idle capacity and is measured in barrels per calendar day or barrels per stream day. **2**. The component of operable capacity is operated at the beginning of the period.

Operating Pressure: Pressure indicated by a gauge when the system is in normal operation (working pressure).

Operation and Maintenance Manual: A manual that describes detailed procedures for operators to follow to operate and maintain specific water or wastewater treatment, pretreatment, or process plants and the equipment of the plants.

Figure 26 A photo of petroleum crude oil refinery.

Operator: **1**. Term used to describe a company appointed by venture stakeholders to take primary responsibility for day-to-day operations for a specific plant or activity. **2**. The company or individual responsible for managing an exploration, development, or production operation. **3**. The company that has the legal authority to drill wells and undertake the production of hydrocarbons that are found. The operator is often part of a consortium and acts on behalf of this consortium. **4**. The company that makes the decisions and is responsible for drilling, completing, operating, and repairing the well.

Operable Utilization Rate: Represents the utilization of the atmospheric crude oil distillation units. The rate is calculated by dividing the gross input to these units by the operable refining capacity of the units.

Organic Compounds: Compounds that include carbon and hydrogen atoms. Generally, organic compounds can be classified as either aliphatics (straight-chain compounds), cyclic (compounds with ring structures), and combinations of aliphatics and cyclic.

Orifice: An opening in a wall or plate used to control the rate of flow into or out of a tank or pipe.

Orifice Meter: A single-phase flow meter, primarily for liquid/gas, that measures the pressure drop created by the hole as gas is flowed (See Figure 27).

Other Hydrocarbons: Materials received by a refinery and consumed as a raw material. Includes hydrogen, coal tar derivatives, gilsonite, and natural gas received by the refinery for reforming into hydrogen. Natural gas to be used as fuel is excluded.

OSHA: **1**. Occupational Safety and Health Administration: US government agency. **2**. The Williams–Steiger Occupational Safety and Health Act of 1970 (OSHA) is a federal law designed to protect the health and safety of industrial workers, including the operators of water supply and treatment systems and wastewater treatment plants. The Act regulates the design, construction, operation, and maintenance of water supply systems, water treatment plants, wastewater collection systems, and wastewater treatment plants. OSHA also refers to the federal and state agencies that administer the OSHA regulations.

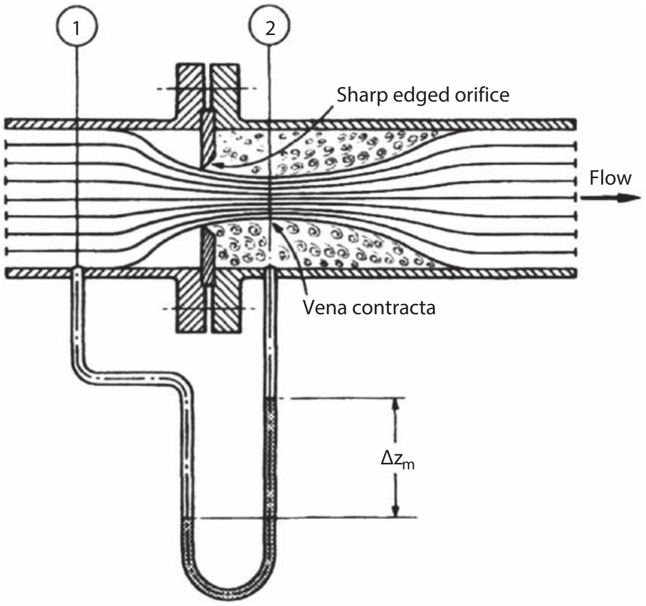


Figure 27 Orifice meter with Vena contracta formation.

Oxidation: Oxidation is the addition of oxygen, the removal of hydrogen, or the removal of electrons from an element or compound. In the environment, organic matter is oxidized to more stable substances. The opposite is reduction.

Oxidation Inhibitor: A substance added in small quantities to a petroleum product to increase its oxidation resistance, thereby lengthening its service or storage life; also called an antioxidant. The oxidation of fuels creates gums that become colloidal, then agglomerate and precipitate. Cracked distillates are found to be more prone to oxidation and deterioration than straight-run distillates. Oxidation fuels can also result in the formation of various acids, ketones, aldehydes, and esters from hydrocarbons. Amino guanidine derivatives when used in the range 3–30 ppm are found effective as antioxidants. Cyclic borates of polymers alkanolamines are effective anti-oxidants even in the 10 ppm range.

Oxidation Stability: **1**. It is used for the evaluation of storage stability and resistance to oxidation as most of the oils, when exposed to air over time, react with oxygen, which are then degraded. Oil with poor oxidation stability, forms corrosive acids at high-temperature conditions in the engine. **2**. Gasoline contains cracked components having a tendency to form gum materials during storage and handling, which affect performance. Oxidation stability provides an indication of the tendency of gasoline and aviation fuels to form gum in storage. In this test, the sample is oxidized inside a stainless-steel pressure vessel initially charged with oxygen at 689 kPa and heated in a boiling water bath. The amount of time required for a specified drop in pressure (gasoline) or the amount of gum and precipitate formed after a specific aging period (aviation fuel) is determined.

Oxidizers: Reactants that oxidize, for example, bleach, chlorine, sodium hypochlorite, and sodium persulfate. Also, a compound that releases oxygen.

Oxidizing Agent: Any substance, such as oxygen (O_2) or chlorine (Cl_2), that will readily add (or take on) electrons. The opposite is a reducing agent.

Oxygen: A chemical element used by all known life forms for respiration.

Oxygenated Fuel: Any organic compound containing oxygen. Specifically for the petroleum industry, this term refers to oxygen-containing organic compounds, such as ethers, and alcohols added to fuels to reduce carbon monoxide in the engine exhausts. They are used as gasoline blending components. Oxygenated fuels tend to give a more complete combustion of its carbon into carbon dioxide (rather than monoxide), thereby reducing air pollution from exhaust emissions.

Oxygenated Fuels Program Reformulated Gasoline: A reformulated gasoline that is intended for use in an oxygenated fuels program control area during an oxygenated fuels program control period.

Oxygenated Gasoline: **1**. Gasoline with an oxygen content of 1.8% or higher by weight that has been formulated for use in motor vehicles. **2**. Finished motor gasoline, other than reformulated gasoline, having an oxygen content 2.7% or higher by weight. It includes gasohol.

Oxygenates: Substances that, when added to gasoline, increase the amount of oxygen in that gasoline blend and thus boost the octane number of gasoline or petrol. Ethanol, methyl tertiary butyl ether (MTBE), ethyl tertiary butyl ether (ETBE), tertiary amyl methyl ether (TAME), and methanol are common oxygenates. MTBE, ETBE, and TAME have 1. Low water solubility, 2. Lower volatility, and 3. A compatibility with hydrocarbon fuels.

Overall Tray Efficiency: Overall tray efficiency can be defined as the number of theoretical trays in a distillation column section divided by the number of actual trays in the section and is reported as a percent. Overall tray efficiencies are less than 100% for all refinery distillation columns.

Overflash: The liquid that returns to the flash zone of a column.

Overhead: Usually refers to the vapor leaving the top tray of a distillation column. For an absorber column, the overhead and the top product are the same.

Overlap: See Gap.

Overpressure: Is any pressure relative to ambient pressure caused by an explosive blast, both positive and negative.

Ozone (O_3): An oxygen molecule with three oxygen atoms that occurs as a blue, harmful, pungent-smelling gas at room temperature. The stratosphere ozone layer, which is a concentration of ozone molecules located at 6–30 mi. above sea level, is in a state of dynamic equilibrium. Ultraviolet radiation not only forms the ozone from oxygen but can also reduce the ozone back to oxygen. The process absorbs most of the ultraviolet radiation from the Sun, shielding life from the harmful effects of radiation.

Packed Bed Scrubber: Vertical or horizontal vessels, partially filled with packing or devices of a large surface area, used for the continuous contact of liquid and gas such that absorption can take place. Frequently, the scrubber liquid or liquor has had chemicals added to react with the absorbed gas.

Packing (Seals): Seals around a moving shaft or other equipment.

Paraffins: **1**. Hydrocarbons of the alkanes family. Paraffins are saturated compounds, i.e., hydrogen atoms are appropriately attached to the carbon atoms such that the carbon atoms have only single bonds in the molecular structure. The general formula for paraffin is C_nH_{2n+2}, where C = carbon atoms, H = hydrogen atoms, and n = 1, 2, 3, 4, 5, …**2**. A white, odorless, tasteless, chemically inert, waxy substance derived from distilling petroleum; a crystalline, flammable substance composed of saturated hydrocarbons. **3**. Normal or straight carbon chain alkanes with carbon chain lengths of C_{18+}. The alkanes in this range solidify at temperatures from 80 to over 200°F (27–93°C).

Partial Pressure: In a gaseous mixture, the pressure contribution for a particular component of the mixture. The sum of the partial pressures of the components in the mixture is the total pressure. For example, in a mixture of two components A and B, with partial pressures as p_A, p_B respectively. The total pressure P_{Total} is: $P_{Total} = p_A + p_B$.

Penetration: A measure of the hardness and consistency of asphalt in terms of the depth that a special pointed device will penetrate the product in a set time and temperature.

Performance Rating: A method of expressing the quality of a high-octane gasoline relative to isooctane. This rating is used for fuels that are of better quality than isooctane.

Petroleum Administration for Defense Districts (PADD): Geographic aggregations of the 50 US states and the District of Columbia into five districts by the Petroleum Administration for Defense in 1950. These districts were originally defined during World War II for purposes of administering oil allocation.

Petroleum Coke: A residue high in carbon content and low in hydrogen that is the final product of thermal decomposition in the condensation process in cracking. This product is reported as marketable coke. The conversion is 5 bbl (of 42 US gal. each) per short ton. Coke from petroleum has a heating value of 6.024 million Btu per barrel.

Marketable coke: Those grades of coke produced in delayed or fluid cokers, which may be recovered as relatively pure carbon. This "green" coke may be sold as is or further purified by calcining.

Catalyst coke: The only catalytic coke used as a fuel is the coke on the catalyst in the FCC process. In other catalytic processes, there is coke deposited on the catalyst, but it is not regenerated in a way such that the heat of combustion is recovered.

Petrolatum: Microcrystalline wax or petroleum jelly.

Petroleum ether: A volatile fraction of petroleum consisting mainly of pentanes and hexanes.

Petrochemical Feedstocks: Chemical feedstocks derived from petroleum principally for the manufacture of chemicals, synthetic rubber, and a variety of plastics. These categories reported are "Naphthas less than 401°F and Other Oils Equal to or greater than 401°F."

Petroleum Products: Petroleum products are obtained from the processing of crude oil (including lease condensate), natural gas, and other hydrocarbon compounds. Petroleum products include unfinished oils, liquefied petroleum gases, pentanes plus, aviation gasoline, motor gasoline, naphtha-type jet fuel, kerosene-type jet fuel, kerosene, distillate fuel oil, residual fuel oil, petrochemical feedstocks, special naphthas, lubricants, waxes, petroleum coke, asphalt, road oil, still gas, and miscellaneous products.

Phase Envelope: **1**. The boundaries of an area on the P-T diagram for the material that encloses the region where both vapor and liquid coexist. **2**. Phase diagram or phase envelope is a relation between temperature and pressure that shows the condition of equilibria between the different phases of chemical compounds, the mixture of compounds, and solutions. Phase diagram is an important issue in chemical thermodynamics and hydrocarbon reservoir. It is very useful for process simulation, hydrocarbon reactor design, and petroleum engineering studies. It is constructed from the bubble line, dew line, and critical point. Bubble line and dew line are composed of bubble points and dew points, respectively. Bubble point is the first point at which the gas is formed when a liquid is heated. Meanwhile, dew point is the first point where the liquid is formed when the gas is cooled. The critical point is the point where all the properties of gases and liquids are equal, such as temperature, pressure, the amount of substance, and others. The critical point is very useful in fuel processing and the dissolution of certain chemicals.

According to the thermodynamic definition of the phase diagram, (phase envelope) is a graph showing the pressure at which the transition of different phases from a compound with respect to temperature. A bubble point that forms a bubble line is a point separating the liquid phase and the two-phase region, namely, the liquid phase and gaseous phase. The dew point that forms the dew line is a point separating the gaseous phase and two-phase region, namely, the liquid and gaseous phase. At the dew point, the following conditions must be satisfied (See Figure 28).

Physical Solvent: A liquid capable of absorbing selected gas components by solubility alone without associated chemical reactions.

Pig: 1. A cylindrical device that is inserted into a pipeline to clean the pipeline wall or monitor the internal condition of the pipeline. **2**. A device for cleaning a pipeline or separating two liquids being removed down the pipeline (intelligent pig—fitted with sensors to check for corrosion or defects in pipelines.). **3**. A flow line clearing device pumped through the line with normal flow. **4**. Refers to a poly pig, which is a bullet-shaped device made of hard rubber or similar material. This device is used to clean pipes. It is inserted in one end of a pipe, moves through the pipe under pressure, and is removed from the other end of the pipe.

Pinch Analysis: Bodo Linnhoff at the University of Leeds in 1977 developed a technique for minimizing energy usage in a process. It is based on calculating the minimum energy consumption by optimizing the heat recovery, energy supply, and process operating conditions. It uses process data represented as energy flows or streams as a function of heat load against temperature. These data are combined for all the hot and cold streams requiring heat. The point of closest approach between the hot and cold composite curves is called the pinch point and corresponds to the point where the design is most constrained. Using this point, the energy targets can be achieved using heat

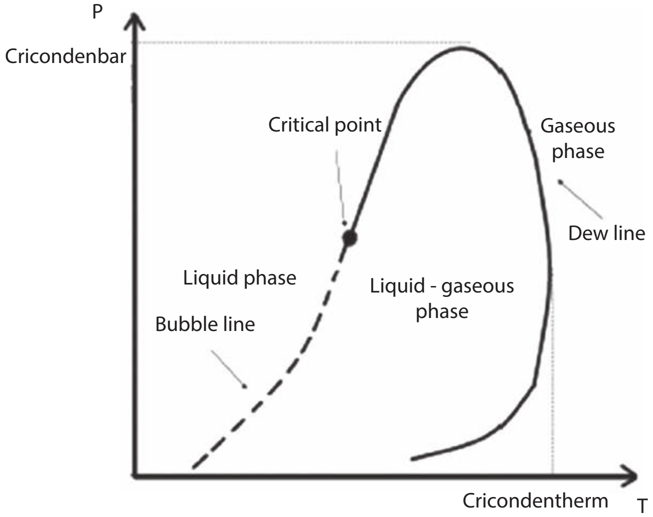


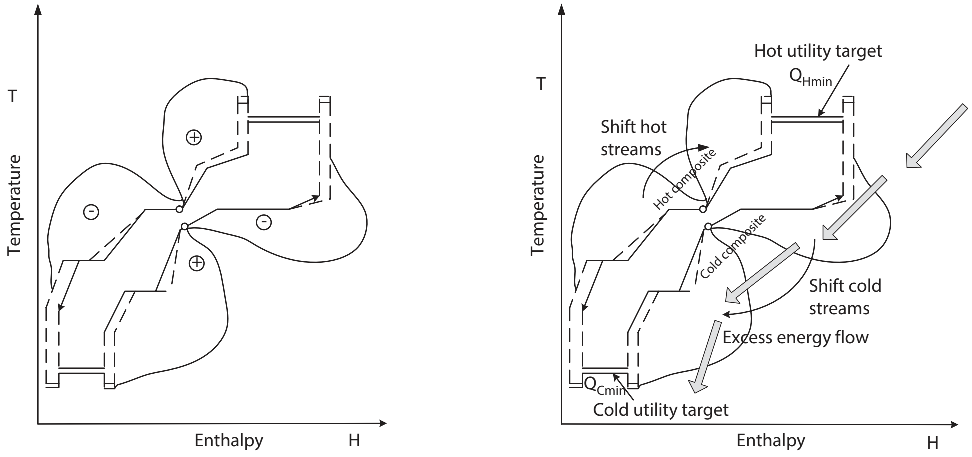
Figure 28 Phase diagram (Phase Envelope).

exchange to recover heat between the hot and cold streams in two separate systems, with one temperature above the pinch temperature and one for the temperature below the pinch temperatures.

Figure 29 shows the point in a pinch analysis that corresponds to the point where the hot and cold streams in an integrated process are most constrained.

Pipelines: Tubular arrangement for the transportation of crude oil, refined products, and natural gas from the well head, refinery, and storage facility to the consumer. The pipeline measures 14–42 in. (356–1067 mm) in diameter but is usually 20–36 in. (508–914 mm). It is often composed of 40 ft. (12 m) lengths, but the lengths may be as long as 60 or 80 ft. (18–24 m). The pipe is wrapped and coated for protection against corrosion, especially since it runs underground. About half of all gases and oils are moved by the pipeline.

Pipe Still: A heater or furnace containing tubes through which oil is pumped while being heated or vaporized. Pipe stills are fired with waste gas, natural gas, or heavy oils, and by providing for rapid heating under conditions of high pressure and temperature, are useful for thermal cracking as well as distillation operations.



(**a**) **The plus-minus principle.**

(**b**) **Shifting streams through the pinch in the right direction enacts the plus-minus principle.**

Figure 29 Shows the point in a pinch analysis that corresponds to the point where the hot and cold streams in an integrated process are most constrained.

Pipe size: Process piping comes in particular nominal sizes:

0.75 in.	-
1 in.	25 mm
2 in.	50 mm
2.5 in.	-
3 in.	80 mm
4 in.	100 mm
6 in.	150 mm
8 in.	200 mm
10 in.	250 mm

The nominal size does not refer either to the outside or inside diameter of the pipe. Pipe thickness affects the ID.

Piping and instrumentation diagram (P & ID): A schematic representation of the interconnecting pipelines and control systems for a process or part of a process (see Figure 30). Uses a standard set of symbols for process equipment and controllers. It includes the layout of branches, reducers, valves, equipment, instrumentation, and control interlocks. It also includes process equipment names and numbers; process piping including sizes and identification; valves and their identification; flow directions, instrumentation, and designations; vents, drains, sampling lines, and flush lines. P & IDs are used to operate the process system, operators' trainings as well as being used in plant maintenance and process modifications. At the design stage, it is useful in carrying out safety and operations investigations such as Hazop. The list of P & ID items are:

- Instrumentation and designations
- Mechanical equipment with names and numbers
- All valves and their identifications
- Process piping, sizes, and identification
- Miscellanea—vents, drains, special fittings, sampling lines, reducers, enlargers, and swagers
- Permanent start-up and flush lines
- Flow directions.
- Interconnections references
- Control inputs and outputs and interlocks
- Interfaces for class changes
- Computer control systems
- Identification of components and subsystems

Plug Flow Reactor (PFR)/Continuous Tubular Reactor (CTR): Is a reactor tubular reactor where fluids enter continuously in an axial direction in a tube as the reaction takes place, and the products are withdrawn at the outlet. The fluid going through a plug flow reactor is modeled as flowing through the reactor as a series of infinitely thin coherent "plugs," each having a uniform composition. The plugs travel in the axial direction of the reactor, with each plug having a different composition from the ones before and after it. The key assumption is that as a plug flows through a PFR, the fluid is perfectly mixed in the radial direction but not mixed at all in the axial direction. Each plug is considered as a separate entity (i.e., effectively an infinitesimally small batch reactor with mixing approaching zero volume). As the plug flows downs the PRF, the residence time of the plug element is derived from its position in the reactor. In this description of the ideal plug flow reactor, the residence time distribution is therefore an impulse (a small narrow spike function). Although it is a powerful tool for estimating purposes, caution is required as a real

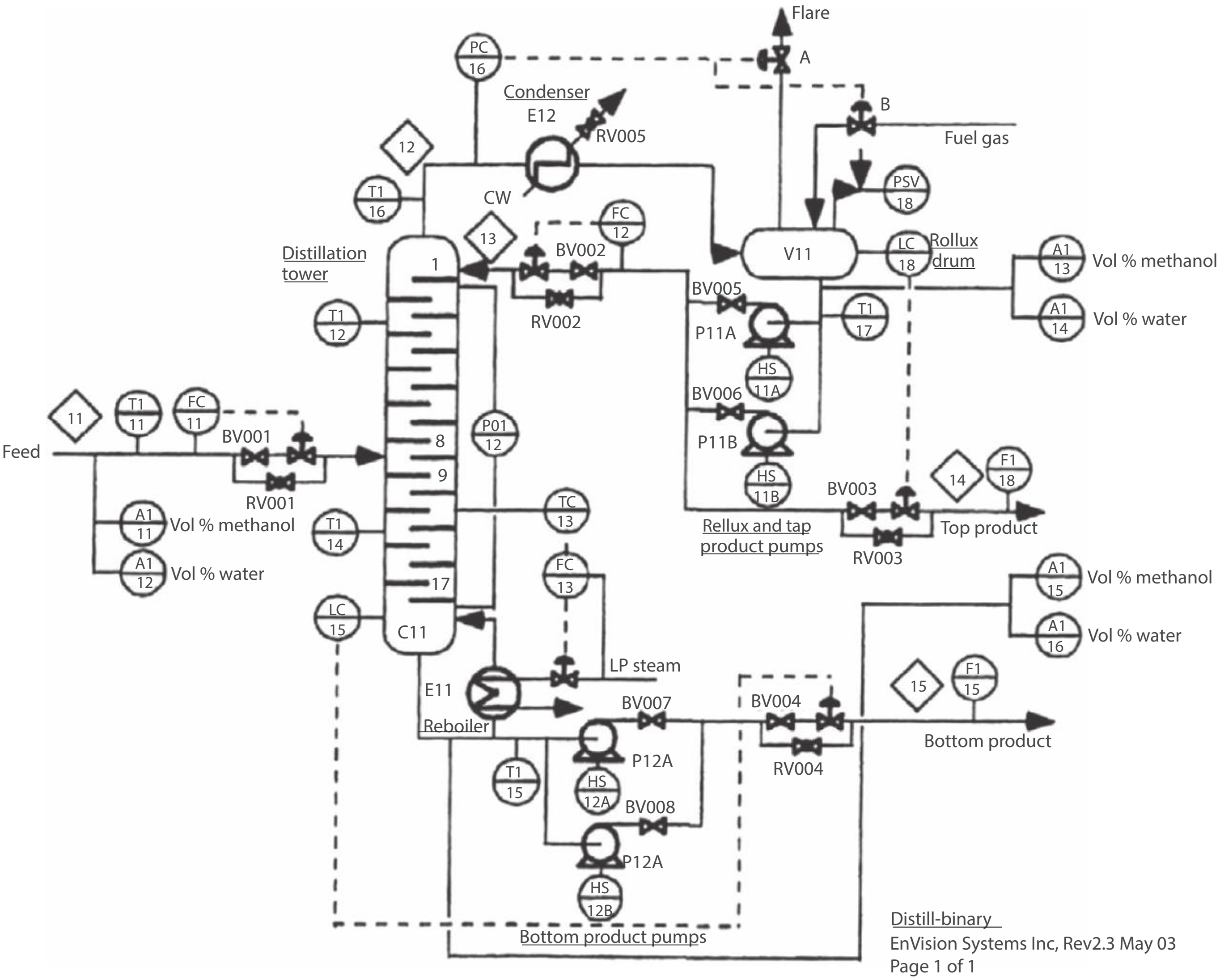


Figure 30 Piping and instrumentation diagram.

flow system exhibits significant variability in residence times. Residence time distribution is one of the factors that need to be considered when scaling flow reactors (Figure 31).

Polymerization: A reaction in which like molecules are joined together to form dimer and trimer compounds, etc., of the reactant(s). This most often occurs with olefinic compounds in oil refineries. The objective of a polymerization unit is to combine or polymerize the light olefins propylene and butylenes into molecules two or three times their original molecular weight. The feed to this process consists of light gaseous hydrocarbons (C_3 and C_4) produced by catalytic cracking, which are highly unsaturated. The polymer gasoline produced has octane numbers above 90.

PONA Analysis: Analysis for paraffins (P), olefins (O), naphthenes (N), and aromatics (A). The method used is ASTM D 1319.

Pour Point: **1**. Is a measure of how easy or difficult to pump the crude oil, especially in cold weather. Specifically, the pour point is the lowest temperature at which a crude oil will flow or pour when it is chilled without disturbance at a controlled rate. The pour point of the whole crude or oil fractions boiling above 450°F (232°C) is determined by the standard test ASTM D97. Both pour and cloud points are important properties of the product streams as far as heavier products are concerned. For heavier products, they are specified in the desired range, and this is achieved

Figure 31 Plug flow reactor, 243' long, 8" S 40 CP pipe. A series of plug flow reactors.

by blending appropriate amounts of lighter intermediate products. **2**. The temperature at which oil starts to solidify and no longer flows freely. Pour point usually occurs 40–42°F (4.5–5.5°C) below the cloud points. A sample tube of petroleum oil is chilled in the pour point test. The pour point is defined as the temperature at which the sample will still pour (move) when the sample tube is tipped. The pour temperature is typically about 5°F (2.8°C) lower than the cloud point.

Power Stroke: Is the downward motion of a piston that occurs after ignition as the fuel combusts and expands.

ppmv: A volume concentration of a species in bulk.

Prandtl number (Pr): A dimensionless number; Pr represents the ratio of the momentum of diffusivity to thermal diffusivity in fluid convection:

$$\mathrm{Pr} = \frac{C_P \mu}{k}$$

where C_p is the specific heat, μ is the viscosity, and k is the thermal conductivity.

Pre-ignition: Describes the event when the air/fuel mixture in the cylinder ignites before the spark plug fires. Pre-ignition is initiated by an ignition source other than the spark, such as hot spots in the combustion chamber, a spark plug that runs too hot for the application, or carbonaceous deposits in the combustion chamber heated to incandescence by previous engine combustion events. It is a technically different phenomenon from engine knocking.

The phenomenon is also referred to as "after-run" or "run-on" or sometimes diesel in when it causes the engine to carry on running after the ignition is shut off. This effect is more readily achieved on carbureted gasoline engines because the fuel supply to the carburetor is typically regulated by a passive mechanical float valve and fuel delivery can feasibly continue until fuel line pressure has been relieved, provided that the fuel can be somehow drawn past the throttle plate.

Pre-ignition and engine knock both sharply increase combustion chamber temperatures. Consequently, either effect increases the likelihood of the other effect occurring, and both can produce similar effects from the operator's perspective, such as rough engine operation or the loss of performance due to operational intervention by a computer. See Knocking.

Pre-Startup Safety Review (PSSR): Audit check performed prior to equipment operation to ensure adequate process safety management (PSM) activities have been performed. The check should verify (1) Construction and equipment is satisfactory, (2) Procedures are available and adequate, (3) A process hazard analysis (PHA) has been undertaken and recommendations resolved, and (4) The employees are trained.

Precursor: Compounds that are suitable or susceptible to specific conversion to another compound, e.g., methyl cyclopentane is a good precursor for making benzene in a catalytic reformer.

Preheat, Preheat Train: Heat exchanger or a network of heat exchangers in which the feed to a process (usually a distillation column) is heated by recovering heat from products being cooled.

Pressure, Absolute: **1**. The force applied over a given area. The instrument gauges used to measure the pressure of fluids are either expressed as absolute pressure, which is measured above a vacuum, or **2**. Gauge pressure plus barometric or atmospheric pressure. The absolute pressure can be zero only in a perfect vacuum. **3**. The pressure is due to the weight of the atmosphere (air and water vapor) on the Earth's surface. The average atmospheric pressure at sea level has been defined as 16.69 lb_f/in^2 absolute.

Pressure, Atmospheric: **1.** The pressure due to the weight of the atmosphere (air and water vapor) on the Earth's surface. The average atmospheric pressure at sea level is 14.696 lb_f/in^2. Absolute. **2**. The pressure exerted by the atmosphere on a given point. It decreases as the elevation above sea level increases.

Pressuring Agent: The hydrocarbon, usually butane, used to bring gasoline blends up to acceptable vapor pressure.

Pressure drop: **1**. The decrease in pressure between two points in a system caused by frictional losses of moving fluid in a pipe or duct or by some other resistance such as across a packed bed, filter, or catalyst, or due to the effects of hydrostatic head such as across the liquid on the tray of a distillation column. **2**. Change in pressure with depth.

Pressure drop multiplier (φ^2): A parameter used in two-phase gas–liquid frictional pressure drop calculations where the overall pressure drop along a length of pipe is due to a combination from the flowing gas and liquid. This is expressed by:

$$\frac{dp_f}{dz} = \varphi_g^2\left(\frac{dp_g}{dz}\right)_g = \varphi_L^2\left(\frac{dp_L}{dz}\right)_L$$

where φ_g^2 and φ_L^2 are the pressure drop multipliers for the liquid and gas phases in which the parameter X^2 is defined as:

$$X = \sqrt{\left[\frac{\left(\frac{dp_L}{dz}\right)_L}{\left(\frac{dp_g}{dz}\right)_g}\right]} = \left(\frac{\varphi_g^2}{\varphi_L^2}\right)^{0.5}$$

Correlations have been developed to determine relationships for the multipliers for combinations of laminar and turbulent gas and liquid phases (See Figure 32).

Pressure, Hydrostatic: The pressure, volume per unit area, exerted by a body of water at rest.

Pressure Integrity Test: A pressure test of a vessel formed by the entire well or part of a well. It usually measures the ability of a pressure vessel to hold pressure without leaking at a given pressure.

Pressure, Negative: A pressure less than atmospheric.

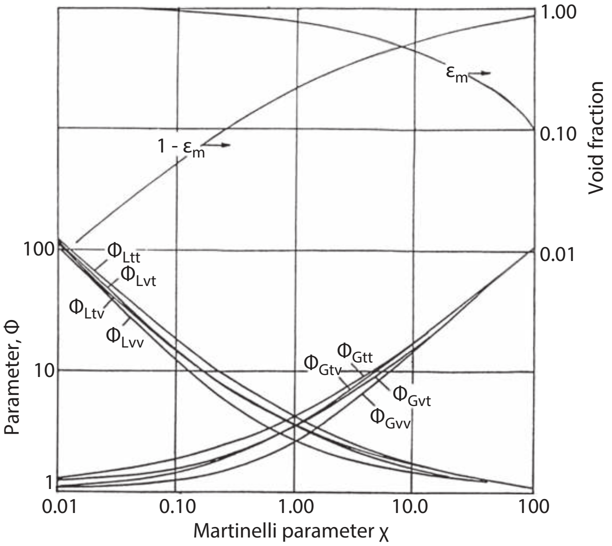


Figure 32 Lockhart–Martinelli two-phase multiplier.

Pressure Reducing Valve: Valve used to reduce high supply pressure to a usable level.

Pressure Relief Valve: A mechanical valve that opens at a preset pressure to relieve pressure in a vessel (See Figure 33).

Primary Absorber: The first absorber in an FCC gas plant.

Pretreatment: A group of processes that natural gas is subjected to prior to its liquefaction. Its purpose is to remove mainstream contaminants or compounds that may cause operational problems in the liquefaction unit.

Pretreatment Facility: Industrial wastewater treatment plant consisting of one or more treatment devices designed to remove sufficient pollutants from wastewaters to allow an industry to comply with effluent limits established by the US EPA General and Categorical Pretreatment Regulations or locally derived prohibited discharge requirements and local effluent limits.

Preventative Maintenance: Maintenance carried out prior to unit or system failure.

Preventive Maintenance: Regularly scheduled servicing of machinery or other equipment using appropriate tools, tests, and lubricants. This type of maintenance can prolong the useful life of equipment and machinery and increase its efficiency by detecting and correcting problems before they cause a breakdown of the equipment.

Probability: The likelihood that the impact or event will occur. Impact (or consequence) is the effect on conditions or people if the hazard is realized (occurs) in practice, and probability is the likelihood that the impact will occur. Risk is a function of probability and impact (consequence). With these discrete data, it is determined by taking the number of occurrences for the particular type of event being considered and dividing that by the total number of outcomes for the event. Expressed as a deterministic value (quantitative single value or high, medium, low, etc.) or as a range of values—that is, uncertainty—that is represented by a probability distribution.

Probability Distribution (Risk): A mathematical relationship between the values and the associated probabilities for a variable across the entire range of possible values for that variable. Typically, probability distributions are displayed as a frequency or cumulative frequency plots.

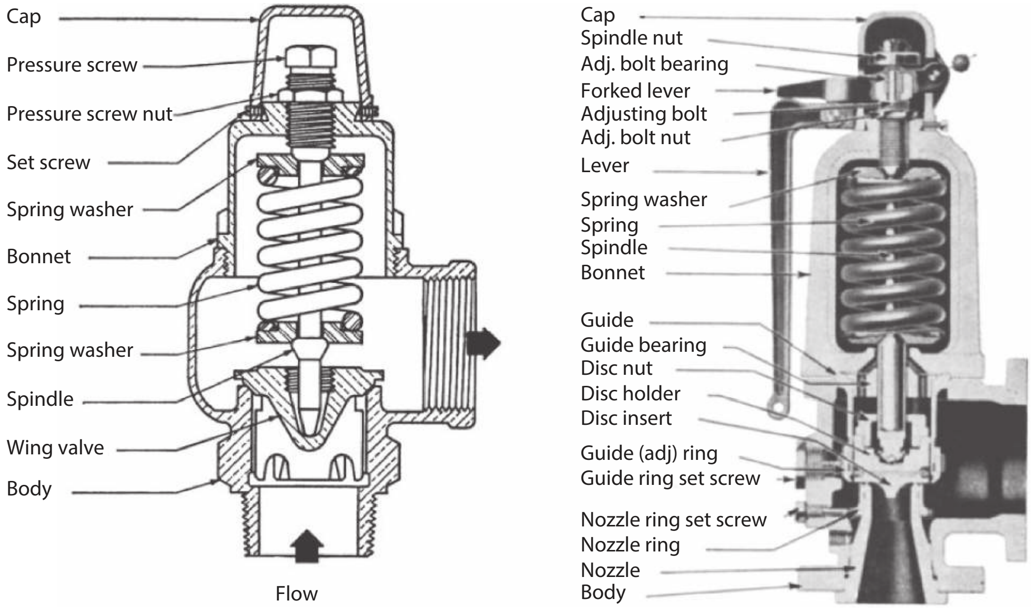


Figure 33 Relief valve and safety valve.

Probability Distillation: The characteristic shape of laboratory distillation boiling curves tends to follow the shape of a normal distribution function, especially the TBP method. A probability distillation paper is constructed with a probability scale for the boiling point scale, and laboratory distillation curves may be plotted as straight lines on this paper. This provides a reasonable way to extrapolate partial laboratory distillation data.

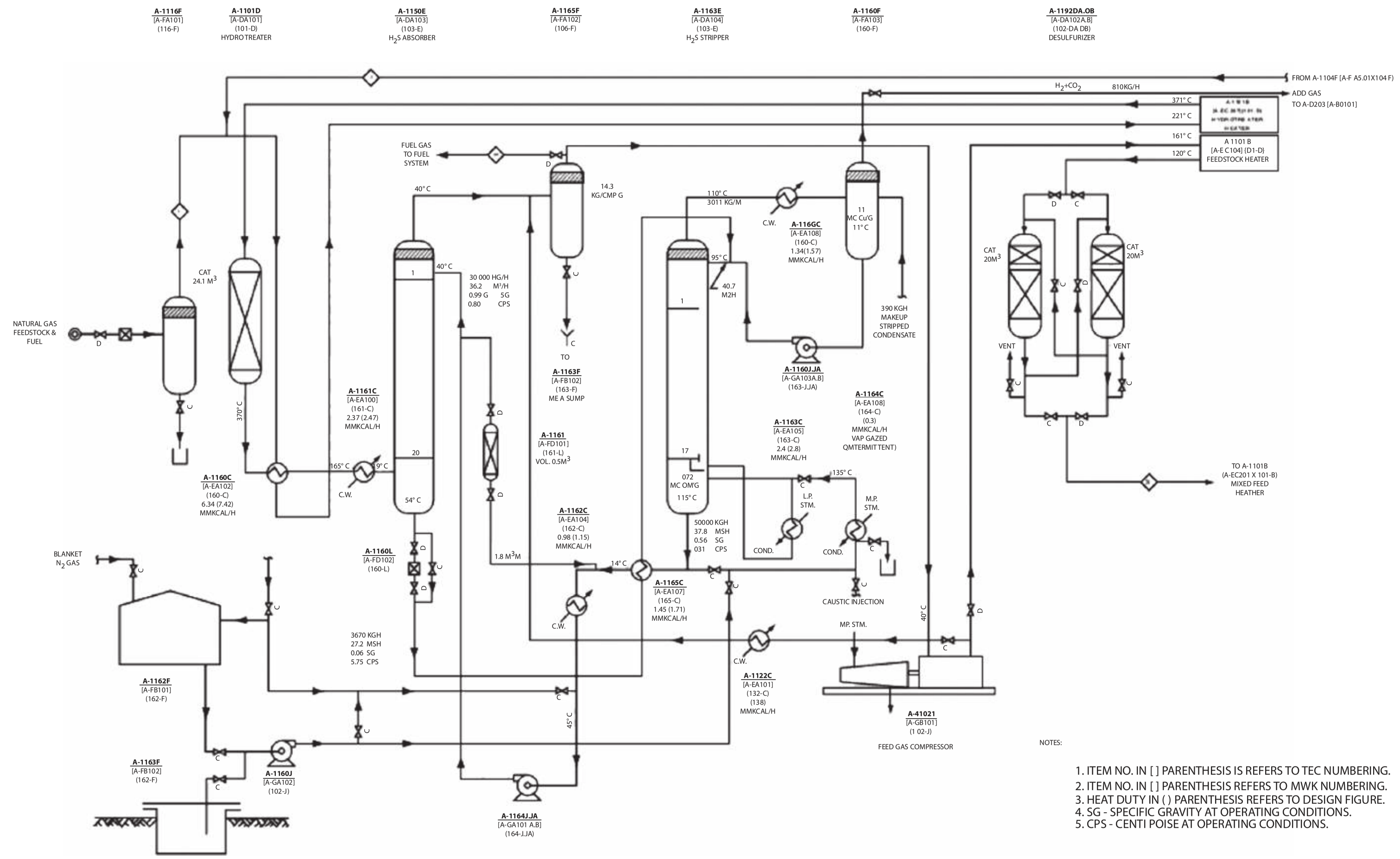
Process: Any activity or operation leading to a particular event.

Process Flow Diagram (PFD): A schematic representation of a process or part of a process that converts raw materials to products through the various units' operations (Figure 34). It typically uses a symbolic representation for the major items of equipment such as storage vessels, reactors, separators, process piping to and from the equipment, as well as by-pass and recirculation lines, and the principal flow routes. Key temperatures and pressures corresponding to normal operation are included, as well as equipment ratings and minimum and maximum operational values. Material flows and compositions are included. It may also include important aspects of control and pumping, as well as any interaction with other process equipment or flows. The design duties or sizes of all the major equipment are also featured, which can collectively provide a comprehensive representation of the process. PFDs generally do not include the following:

- Pipe classes or piping line numbers
- Process control instrumentation (sensors and final elements)
- Minor bypass lines
- Isolation and shutoff valves
- Maintenance vents and drains
- Relief and safety valves
- Flanges

Programmable Logic Controller (PLC): A digital electronic controller that uses a computer-based programmable memory for implementing operating instructions through digital or analog inputs and outputs.

Process Hazard Analysis (PHA): An organized formal review to identify and evaluate hazards with industrial facilities and operations to enable their safe management. The review normally employs a qualitative technique to identify and access the importance of hazards as a result of identified consequences and risks. Conclusions and recommendations are provided for risks that are deemed at a level that is not acceptable to the organization. Quantitative methods may be also employed to embellish the understanding of the consequences and risks that have been identified.



NOTES:

1. ITEM NO. IN [ ] PARENTHESIS IS REFERS TO TEC NUMBERING.
2. ITEM NO. IN [ ] PARENTHESIS REFERS TO MWK NUMBERING.
3. HEAT DUTY IN ( ) PARENTHESIS REFERS TO DESIGN FIGURE.
4. SG - SPECIFIC GRAVITY AT OPERATING CONDITIONS.
5. CPS - CENTI POISE AT OPERATING CONDITIONS.

Figure 34 Process flow diagram (feed and fuel desulfurization section).

Process Risk: Risk arising from the process conditions caused by abnormal events (including basic process control system (BPCS) malfunction).

Note: The risk in this context is that associated with the specific hazardous event in which Safety Instrument Systems (SIS) are to be used to provide the necessary risk reduction (i.e., the risk associated with functional safety).

Process Safety Management (PSM): A comprehensive set of plans, policies, procedures, practices, administrative, engineering, and operating controls designed to ensure that barriers to major incidents are in place, in use, and are effective.

Processing Gain: The volumetric amount by which the total output is greater than the input for a given period of time. This difference is due to the processing of crude oil into products that in total have lower specific gravity than the crude oil processed.

Processing Loss: The volumetric amount by which the total refinery output is less than the input for a given period of time. This difference is due to the process of crude oil into products that in total have higher specific gravity than the crude oil processed.

Production Capacity: The maximum amount of product that can be produced from processing facilities.

Products Supplied: **1**. Crude Oil: Crude oil burned on leases and by pipelines as fuel. **2**. Approximately represents the consumption of petroleum products because it measures the disappearance of these products from primary sources, i.e., refineries, natural gas processing plants, blending plants, pipelines, and bulk terminals. In general, the product supplied of each product in any given period is computed as follows: field production, plus refinery production, plus imports, plus unaccounted for crude oil (plus net receipts when calculated on a PAD District basis), minus stock change, minus crude oil losses, minus refinery inputs, minus exports.

Propane (C_3H_8): A hydrocarbon gas that is a principal constituent of the heating fuel, LPG. Propane is used extensively for domestic heating and as a feed to ethylene plants.

Propylene (C_3H_6): A hydrocarbon in the olefin series resulting from olefin plant operations and refinery cracking processes and used as an alkyl plant feed or chemical feedstock.

Propylene (C_3H_6) (non-fuel use): Propylene that is intended for use in non-fuel applications such as petrochemical manufacturing. Non-fuel use propylene includes chemical-grade propylene, polymer-grade propylene, and trace amounts of propane. Non-fuel use propylene also includes the propylene component of propane/propylene mixes where the propylene will be separated from the mix in a propane/propylene splitting process. Excluded is the propylene component of propane/propylene mixes where the propylene component of the mix is intended for sale into the fuel market.

Process design: The design of an industrial process that uses physical, chemical, or biochemical transformations for the production of useful products. It is used for the design of new processes, plant modifications, and revamps (Figure 35). It starts with conceptual and feasibility studies and includes detailed material and energy balances, the production of block flow diagrams (BFDs), process flow diagrams (PFDs), engineering line diagrams (ELDs), and piping and instrumentation diagrams (P & IDs). It also includes the production of reports and documents for plant construction, commissioning, start-up, operation, and shutdown. The reports and documents are used by vendors, regulatory bodies, operators, and other engineering disciplines.

Process economics: An evaluation of a process in terms of all the costs that are involved. It considers the cost of raw materials and how they are processed, as well as the costs associated with waste processing such as recycling or disposal. It also includes the optimization of a process to best utilize materials and energy. The fixed costs of a process are not dependent on the rate of production, but the variable costs are and must be met by the revenue generated by sales. Taxes are deducted, resulting in the net profit.

Process engineer: He/she uses the principles of heat and material balances, hydraulics, vapor–liquid equilibrium, and chemistry to solve plant operating problems and optimize operating variables.

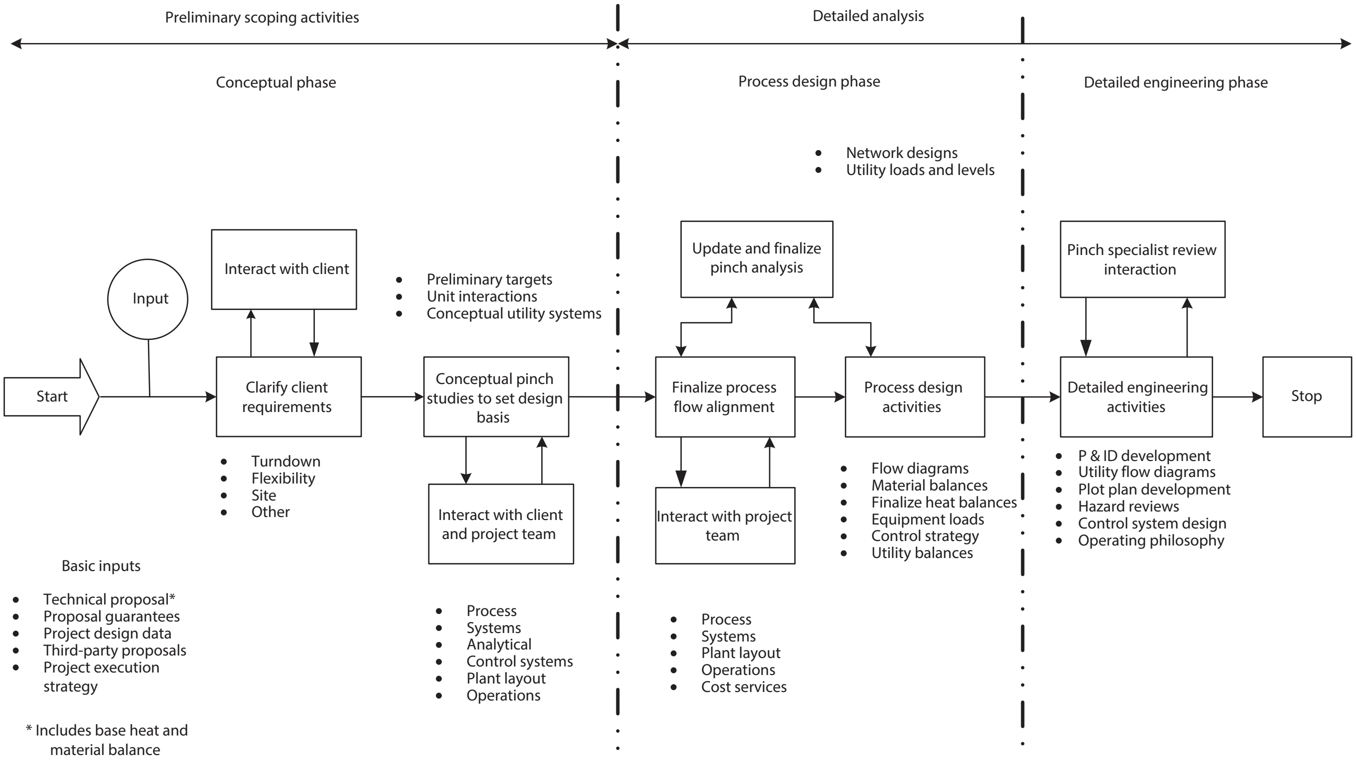


Figure 35 This new process design work process implements process integration effectively.

Process Integration: **1**. A holistic approach used in process design that considers the process as a whole with the interactions between unit operations in comparison with the optimization of unit operations separately and independently. It is known as process synthesis (See Figure 36). **2**. A technique used to minimize the energy consumption and heat recovery in a process. It is also known as process heat integration and pinch analysis (See Energy Management).

Process Intensification: An approach to the engineering design, manufacture, and operation of processes that aims to substantially improve process performance through energy efficiency, cost-effectiveness, reduction in waste, improvement in purification steps, reduction of equipment size, and increase in safety and operational simplicity. It involves a wide range of innovative reactor, mixing, and separation technologies that can result in dramatic improvements in process performance. It involves an integrative approach that considers overall process objectives rather than the separate performance of individual unit operations; process intensification can enable a process to achieve its maximal performance, leading to the development of cheaper, smaller, cleaner, safer, and sustainable technologies.

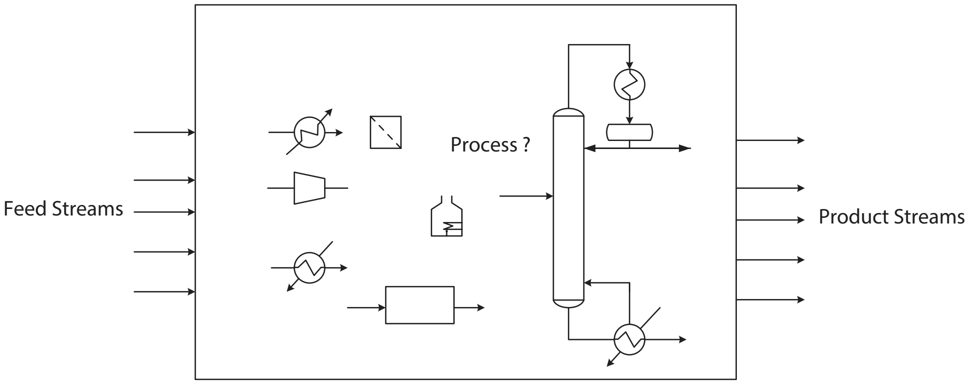


Figure 36 Process integration starts with the synthesis of a process to convert raw materials into desired products.

Any chemical engineering development that leads to a substantially smaller, cleaner, safer, sustainable, and more energy-efficient technology. Process Intensification (PI) can be employed in inherently safer plants and is grouped into four major strategies:

- Minimize
- Substitute
- Moderate
- Simplify

PI can allow one to moderate conditions to minimize the risk of explosions and to simplify processes by having fewer unit operations and less complex plants.

Process plant: A collective name for an industrial facility used to convert raw materials into useful products. It includes all the process equipment such as reactors, mixers and separating units, all the associated pipework and pumps, heat exchangers, and utilities such as steam and cooling water.

Process Safety: A comprehensive management system that focuses on the management and control of potential major hazards that arise from process operations. It aims at reducing risk to a level that is as low as is reasonably practicable by the prevention of fires, explosions, and accidental or unintended chemical releases that can cause harm to human life and to the environment. It includes the prevention of leaks, spills, equipment failure, over- and under-pressurization, over-temperatures, corrosion, and metal fatigue. It covers a range of tools and techniques required to ensure a safe operation of the plant and machinery to ensure the safety of personnel, the environment, and others through detailed design and engineering facilities, the maintenance of equipment, the use of effective alarms and control points, procedures, and training. It also includes risk assessment, layers of protection analysis, and the use of permit-to-work authorizations.

Process Simulation: The use of computers to model and predict the operational and thermodynamic behaviors of a process. Commercial software packages are used to simulate and model batch, continuous, steady-state, and transient processes. They require combined material and energy balances, the properties of the materials being processed, and sometimes combine the use of experimental data with mathematical descriptions of the process being simulated. Most software packages feature optimization capabilities involving the use of complex cost models and detailed process equipment size models. Some commercial software products are shown in the table below:

Software	Developer	Applications	Website
Aspen Plus/Aspen Hysys	Aspen Technology	Process simulation and optimization	www.aspentech.com
CHEMCAD	Chemstations	Software suite for process simulation	www.chemstations.com
Design II for Windows	WinSim Inc.	Process simulation	www.winsim.com
gPOMS	PSE Ltd.	Advanced process simulation and modeling	www.psenterprise.com
PRO II	SimSci	Process simulation	www.software.schneider-electric.com/simsci
ProSim Plus	ProSim	Process simulation and optimization	www.prosim.net
UniSim	Honeywell	Process simulation and optimization	www.honeywellprocess.com

Process synthesis: The conceptual design of a process that identifies the best process flowsheet structure, such as the conversion of raw materials into a product. This requires the consideration of many alternative designs. The complex structure of most processes is such that the flowsheet is split into smaller parts and each is reviewed in turn. Then, choices and decisions are made. Many techniques are used in arriving at the best flowsheet such as those based on total cost, which needs to be minimized. Use is made of graphical methods, heuristics, and various other forms of minimization such as the use of process integration.

Process upset: A sudden, gradual, or unintended change in the operational behavior of a process. It may be due to process equipment failure or malfunction, operator intervention, a surge or fall in pressure, flow, level, concentration, etc.

Process variable: A dynamic feature of a process or system that is required to be controlled to ensure that it operates according to design requirements and does not deviate to be unsafe or result in undesirable consequences. The commonly measured process variables include temperature, pressure, flow, level, and concentration.

Pseudo-component: For engineering calculation purposes, a component that represents a specified portion of the TBP distillation curve for a petroleum mixture. The pseudo-component is assigned a normal boiling point and gravity corresponding to the average for the boiling point range. The molecular weight and other properties are derived from the boiling point and gravity using literature correlations for hydrocarbons.

Pump: A mechanical device used to transport a fluid from one place or level to another by imparting energy to the fluid. The three bond groupings are centrifugal, reciprocating, and rotary-type pumps. The most commonly used is the centrifugal type, which has a rotating impeller used to increase the velocity of the fluid and where a part of the energy is converted to pressure energy. Rotary and reciprocating pumps are positive displacement pumps in which portions of fluid are moved in the pump between the teeth of gears and by the action of a piston in a cylinder. There are many variations of these types, and each has a particular application and suitability for fluid in terms of its properties, required flow rate, and delivery pressure (See Figure 37).

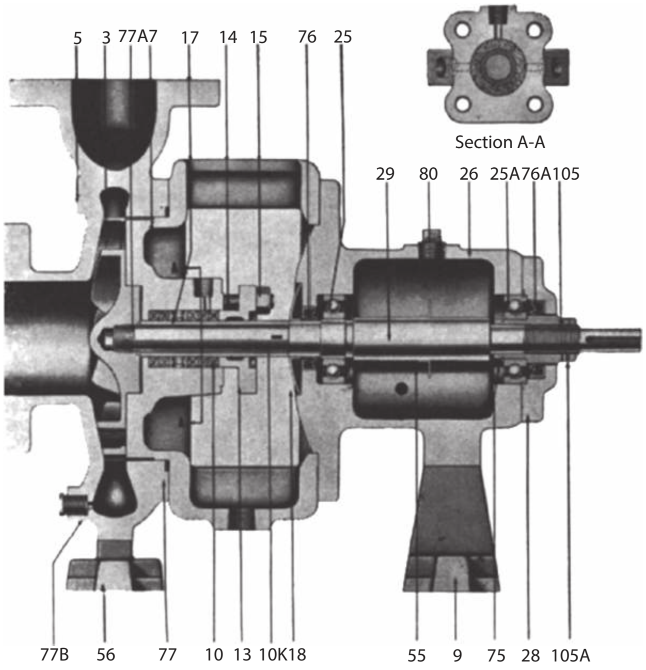
Pumparound: A liquid side-draw from a distillation/fractionating column that is pumped, cooled, and returned to a higher location in the column. Pumparounds recover usable heat that would be lost at the condenser. They also lower the vapor flow in a column and reduce the required column diameter for vapor-loaded columns such as crude and vacuum columns.

Pumpdown: A liquid side draw that is pumped down to a tray below the draw tray, usually the next tray lower. Pumpdowns are sometimes cooled prior to returning to the column.

Pump priming: Used for the start-up and successful operation of centrifugal pumps in which the casing housing the "impeller" is first filled or primed with liquid before operation begins. Since the density of a liquid is many times greater than that of a gas, vapor, or air, the suction pressure is otherwise insufficient to draw in more liquid. Depending on the type of pump, priming can be achieved either manually or by drawing liquid using a vacuum pump. Valves can be used to prevent drainage and ensure that the pump does not require priming once the pump stops.

Purge: A stream that is removed from a recycling process to prevent the buildup of one or more components in the process streams.

Pyrolysis: **1**. Heating a feedstock to high temperature to promote cracking as in an ethylene plant. **2**. Destructive distillation that involves the decomposition of coal, woody materials, petroleum, and so on, by heating in the absence of air.



3	Impeller
5	Casing
7	Back head cradle
9	Bearing housing foot
10	Shaft sleeve
10K	Shaft sleeve key
13	Stuffing box gland
14	Stuffing box gland stud
15	Stuffing box gland stud Nut
17	Seal cage
18	Splash collar
25	Shaft bearing—radial
25A	Shaft bearing—thrust
26	Bearing housing
28	Bearing end cover
29	Pump shaft
55	Oil disc. (flinger)
56	Casing foot
75	Retaining ring
76	Oil seal—front
76A	Oil seal—rear
77	Gasket—casing
77A	Gasket—sleeve
77B	Gasket—drain plug
80	Oil vent
105	Shaft adjusting sleeve
105A	Sleeve lock nut

Figure 37 General service centrifugal pump.

Pyrolysis Gasoline: The gasoline created in an ethylene plant cracking gas oil or naphtha feedstocks. Sometimes called pygas, it has a high content of aromatics and olefins and some diolefins.

Pyrophoric Iron Sulfide: A substance typically formed inside tanks and processing units by the corrosive interaction of sulfur compounds in the hydrocarbons and the iron and steel in the equipment. On exposure to air (oxygen), it ignites spontaneously.

Quality: The weight fraction of vapor in a vapor–liquid mixture.

Quench: Hitting a very hot stream coming out of a reactor with a cooler stream to stop immediately the reaction runaway.

Quench Crack: A crack in the steel resulting from stresses produced during the transformation from austenite to martensite.

Quench Hardening: Heat treating requiring austenitization followed by cooling, under conditions that austenite turns into martensite.

Quenching Oil: An oil introduced into high-temperature process streams during refining to cool them.

Quench Stream: A cooled stream that is used to cool another stream by direct contact. For example, hydrogen quench streams are used to quench the hot effluents from hydrocracker reactors.

Quench Zone: A section of a distillation column where a hot stream, usually vapor, is cooled by direct contact with a stream that has been cooled, usually a liquid.

Radiant Heat Transfer: Heat transfer without convection or conduction. Sunshine is radiant heat.

Radiation: Transmission of energy by means of electromagnetic waves emitted due to temperature.

Radical: A group of atoms that separate themselves from a compound momentarily and are highly reactive. For example, two methyl radicals $^*CH_3$ can come from cracking an ethane compound, but they will rapidly attach themselves to some other atom or compound.

Raffinate: **1**. The leftover from a solvent extraction process. **2**. In solvent refining, that portion of the oil that remains undissolved and is not removed by the selective solvent.

Rating Calculations: Calculations in which a unit operation such as a column, heat exchanger, pump, etc., is checked for capacity restrictions.

Ratio of Specific Heats: **1**. Thermodynamic comparison ($k = C_p/C_v$) of the ratio of specific heat (k) at constant pressure (C_p) to specific heat at constant volume (C_v). The ratio range of most gases is 1.2–1.4. 2. For gases, it is the ratio of the specific heat at constant pressure to the specific heat at constant volume. The ratio is important in thermodynamic equations as compressor horsepower calculations and is given the symbol k, where $k = C_p/C_v$. The ratio lies between 1.2 and 1.4 for most gases.

Reactor: The vessel in which chemical reactions take place (See Figure 38).

Reactive Distillation: A distillation column in which there is a section designed for chemical reaction, usually containing a catalyst bed. Some MTBE and TAME processes use a reactive distillation column in place of a second reactor prior to the product separation column.

Reactor Effluent: The outlet stream from a reactor.

Reboiler: **1**. A heat exchanger used toward the bottom of a fractionator to reheat or oven-vaporize a liquid and introduce it several trays higher to help purify the incoming stream or get more heat into the column. **2**. An auxiliary unit of a fractionating tower designed to supply additional heat to the lower portion of the tower (Figure 39).

Recovery: Usually refers to the fraction expressed as a percent of a component or group of components in the feed to a distillation column that is recovered in a given product stream.

Rectification Zone: The portion of a distillation column in which heavy components are washed down the column by contact with a liquid reflux stream. In conventional distillation columns, this is the portion of the column from the tray above the feed tray to the top tray.

Recycle: A process stream that is returned to an upstream operation.

Recycled Feeds: Streams that have been processed and are fed back to the reactors for additional processing.

Reduced Crude: A residual product remaining after the removal by distillation of an appreciable quantity of the more volatile components of crude oil.

Figure 38 Batch reactor types with (a) constant flux (Coflux) jacket, (b) half coil jacket, and (c) a single external cooling jacket.

Reduced Pressure: The ratio of the absolute pressure to the critical pressure.

Reduced Temperature: The ratio of the absolute temperature to the critical temperature.

Reducing Agent: Any substance, such as base metal (iron) or the sulfide ion, that will readily donate (give up) electrons. The opposite is an oxidizing agent.

Reduction: The addition of hydrogen, the removal of oxygen, or the addition of electrons to an element or compound. Under anaerobic conditions (no dissolved oxygen present), sulfur compounds are reduced to odor-producing hydrogen sulfide (HS) and other compounds.

Redwood Viscometer: Standard British viscometer. The number of seconds required for 50 ml of oil to flow out of a standard Redwood viscometer at a definite temperature is the Redwood viscosity.

Refinery Grade Butane (C_4H_{10}): A refinery-produced stream that is composed predominantly of normal butane and/or isobutane and may also contain propane and/or natural gasoline. These streams may also contain significant levels of olefin and/or fluoride contamination.

Refinery Input, Crude Oil: Total crude oil (domestic plus foreign) input to crude oil distillation units and other refinery processing units (cokers, etc.)

Refined Products: The various hydrocarbons obtained as a result of refining process separation from crude oil. Typical refined products are LPG, naphtha, gasoline, kerosene, jet fuel, home heating oil, diesel fuel, residual fuel oil, lubricants, and petroleum coke.

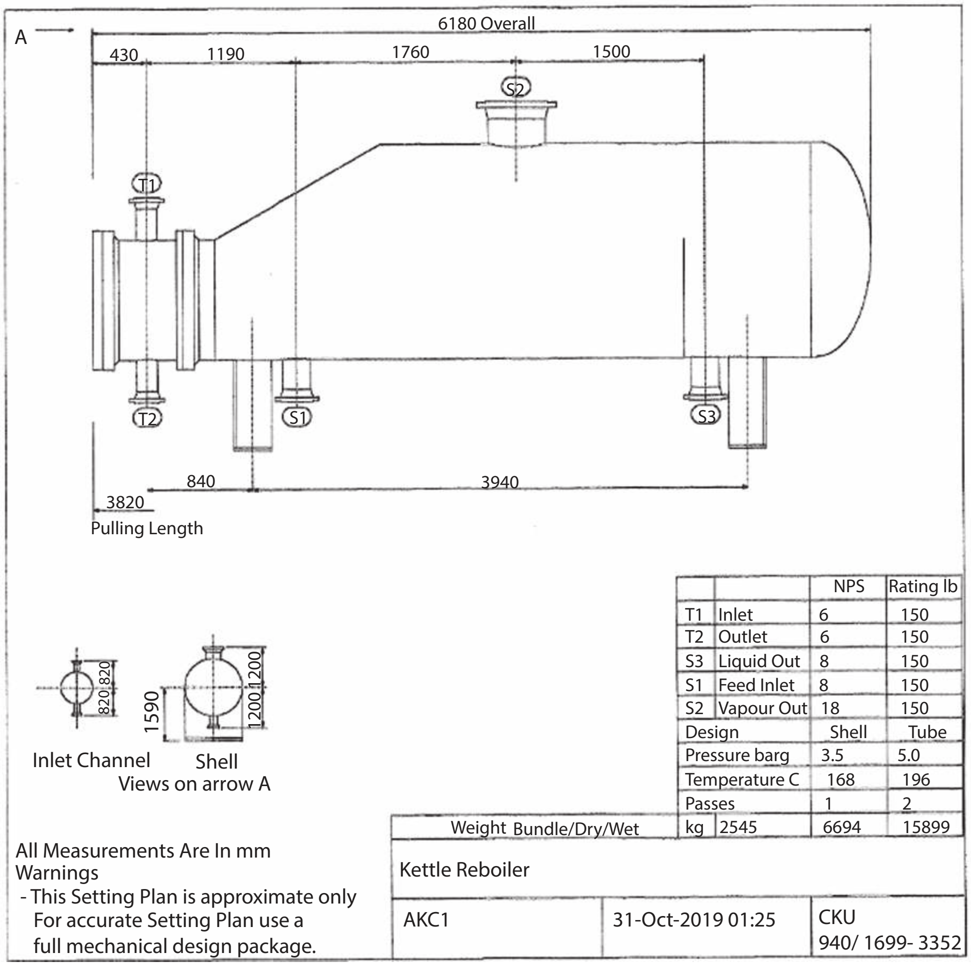


Figure 39 Kettle Reboiler diagram.

Refiner: A company involved in upgrading hydrocarbons to saleable products.

Refinery: **1**. An installation that manufactures finished petroleum products from crude oil, unfinished oils, natural gas liquids, other hydrocarbons, and oxygenates. **2**. A plant used to separate the various components present in crude oil and convert them into usable fuel products or feedstock for other processes. **3**. A large plant composed of many different processing units that are used to convert crude oil into finished or refined products. These processes include heating, fractionating, reforming, cracking, and hydrotreating.

Refinery Gas: A non-condensable gas collected in petroleum refineries.

Refinery Input (Crude Oil): Total crude oil (domestic plus foreign) input to crude oil distillation units and other refinery processing units (cokers).

Refinery Input (total): The raw materials and intermediate materials processed at refineries to produce finished petroleum products. They include crude oil, products of natural gas processing plants, unfinished oils, other hydrocarbons and oxygenates, motor gasoline and aviation gasoline blending components, and finished petroleum products.

Refinery Margin: The difference in value between the products produced by a refinery and the value of the crude oil used to produce them. Refining margins will thus vary from refinery to refinery and depend on the price and characteristics of the crude used.

Refinery Production: Petroleum products produced at a refinery or blending plant. Published production of these products equals refinery production minus refinery input. Negative production occurs when the amount of a product produced during the month is less than the amount of that same product that is reprocessed (input) or reclassified to become another product during the same month. The refinery production of unfinished oils and motor and aviation gasoline blending components appear on a net basis under refinery input.

Refinery Yield: Represents the percentage of finished product produced from the input of crude oil and net input of unfinished oils (expressed as a percentage). It is calculated by dividing the sum of crude oil and net unfinished input into the individual net production of finished products. Before calculating the yield of finished motor gasoline, the input of natural gas liquids, other hydrocarbons and oxygenates, and the net input of motor gasoline blending components must be subtracted from the net production of finished motor gasoline. Before calculating the yield of finished aviation gasoline, the input of aviation gasoline blending components must be subtracted from the net production of the finished aviation gasoline.

Redwood Viscometer: Standard British viscometer. The number of seconds required for 50 ml of oil to flow out of a standard Redwood viscometer at a definite temperature is the Redwood viscosity.

Reflux: 1. Condensed liquid that is returned to the top tray of a distillation column. Reflux helps rectify the mixture being distilled by washing heavy components down the column. **2**. The portion of the distillate returned to the fractionating column to assist in attaining better separation into desired fractions.

Reflux drum: A drum that receives the outlet from the overhead condenser from a distillation column. The liquid and vapor portions are separated in the reflux drum.

Reformate: An upgraded naphtha resulting from catalytic or thermal reforming.

Reforming: **1**. The mild thermal cracking of naphthas to obtain more volatile products such as olefins, of higher-octane values, or the catalytic conversion of naphthas' components to produce higher-octane aromatic compounds. **2**. A refining process used to change the molecular structure of a naphtha feed derived from crude oil by distillation. **3**. The gasoline produced in a catalytic reforming operation.

Reformulated Fuels: Gasoline, diesel, or other fuels that have been modified to reflect environmental concerns, performance standards, government regulations, customer preferences, or new technologies.

Reformed Gasoline: Gasoline made by a reformate process.

Reformulated Gasoline (RFG): **1**. A gasoline whose composition has been changed (from that of gasoline sold in 1990) to (a) include oxygenates, (b) reduce the content of olefins and aromatics and volatile components, and (c) reduce the content of heavy hydrocarbons to meet performance specifications for ozone-forming tendency and for the release of toxic substances (benzene, formaldehyde, acetaldehyde, 1,3-butadiene, and polycyclic organic matter) into the air from both evaporation and tailpipe emissions. **2**. Is a cleaner-burning gasoline that reduces smog and other forms of air pollution. Federal law mandates the sale of reformulated gasoline in metropolitan areas with the worst ozone smog. **3**. Finished motor gasoline formulated for use in motor vehicles, the composition, and properties of which meet the requirements of the reformulated gasoline regulations promulgated by the US Environmental Protection Agency under Section 211 (k) of the Clean Air Act. NB: This category includes the oxygenated fuels program reformulated gasoline (OPRG) but excludes reformulated gasoline blendstock for oxygenate blending (RBOB). **4**. Gasoline that meets the requirements imposed by the Clean Air Act Amendment, passed by the United States Congress on November 15, 1990. Restrictions were placed on volatile organic compounds, nitrous oxides (NO_x) from combustion, and toxins primarily related to benzene (C_6H_6) and its derivatives.

Reformulated Gasoline Blendstock for Oxygenate Blending: A motor gasoline blending component that, when blended with a specified type and percentage of oxygenate, meets the definition of reformulated gasoline.

Refrigerant: **1**. In a refrigerating system, the medium of heat transfer that picks up the heat by evaporating at a low temperature and pressure and gives up the heat on condensing at a higher temperature and pressure. **2**. It is the fluid that performs an inverse thermodynamic cycle, generating the low temperature required for natural gas cooling and liquefaction.

Refrigerant Compressor: A component of a refrigerating system that increases the pressure of a compressible refrigerant fluid and simultaneously reduces its volume while moving the fluid through the device.

Refrigerating System: A system that, in operation between a heat source (evaporator) and a heat sink (condenser), at two different temperatures, is able to absorb heat from the heat source at the lower temperature and reject heat to the heat sink at the higher temperature.

Refrigeration: The process used to remove the natural gas liquids by cooling or refrigerating the natural gas until the liquids are condensed out. The plants use Freon or propane to cool the gas.

Refrigeration (or cooling cycle): **1.** The process used to remove the natural gas liquids by cooling or refrigerating the natural gas until the liquids are condensed out. The plants use Freon or propane to cool the gas. **2.** Inverse thermodynamic cycle whose purpose is to transfer heat from a medium at a low temperature to a medium at a higher temperature.

Regasification: The process by which LNG is heated, converting it into its gaseous state.

Regasification Plant: A plant that accepts deliveries of LNG and vaporizes it back to gaseous form by applying heat so that the gas can be delivered into a pipeline system.

Regenerator: The vessel in a catalytic process where a spent catalyst is cleaned up before being recycled back to the process. An example is the catalytic cracker regenerator where coke is burned off the catalyst.

Regeneration: **1**. The process of burning off coke deposits on catalyst with an oxygen containing gas under carefully controlled conditions. **2**. In a catalytic process, the reactivation of the catalyst, sometimes done by burning off the coke deposits under carefully controlled conditions of temperature and oxygen content of the regeneration gas stream (See Figure 40).

Reid Vapor Pressure (RVP): An ASTM test method to determine the vapor pressure of a light petroleum stream. The Reid vapor pressure is very nearly equal to the true vapor pressure for gasoline streams. There is also a Reid vapor pressure test for crude oil (See Figure 41 and Table 3).

Relative volatility (α): The ratio of the vapor pressure of one liquid component to another in a heterogeneous mixture and a measure of their separability. For a binary mixture, the relative volatility can be expressed in terms of the mole fraction of the more volatile component in the liquid and vapor phases, x and y, as:

$$\alpha = \frac{y(1-x)}{x(1-y)}$$

The greater the value of the relative volatility, the greater the degree of separation. If y = x, then no separation is possible.

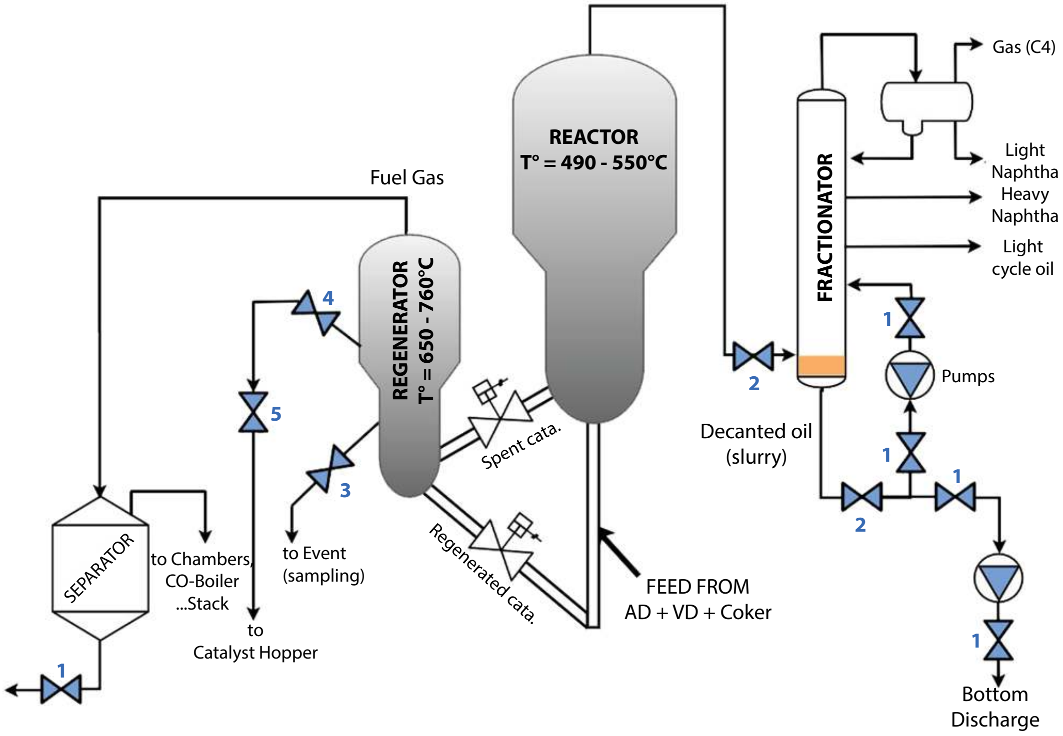


Figure 40 A diagram of an FCC showing the regenerator.

Reliability: The probability that a component or system will perform its defined logic functions under the stated conditions for a defined period of time.

Research Octane Number (RON): One of the two standard tests of gasoline knock; this one simulates less severe operating conditions like cruising. It is determined in a special laboratory test engine under mild "engine-severity" conditions, giving a measure of the low-speed knock properties of gasoline. In contrast with the Motor Octane Number.

Residence time: **1.** The amount of time a hydrocarbon spends in a vessel where a reaction occurs. **2**. The period of time in which a process stream will be contained within a certain volume or piece of equipment, seconds.

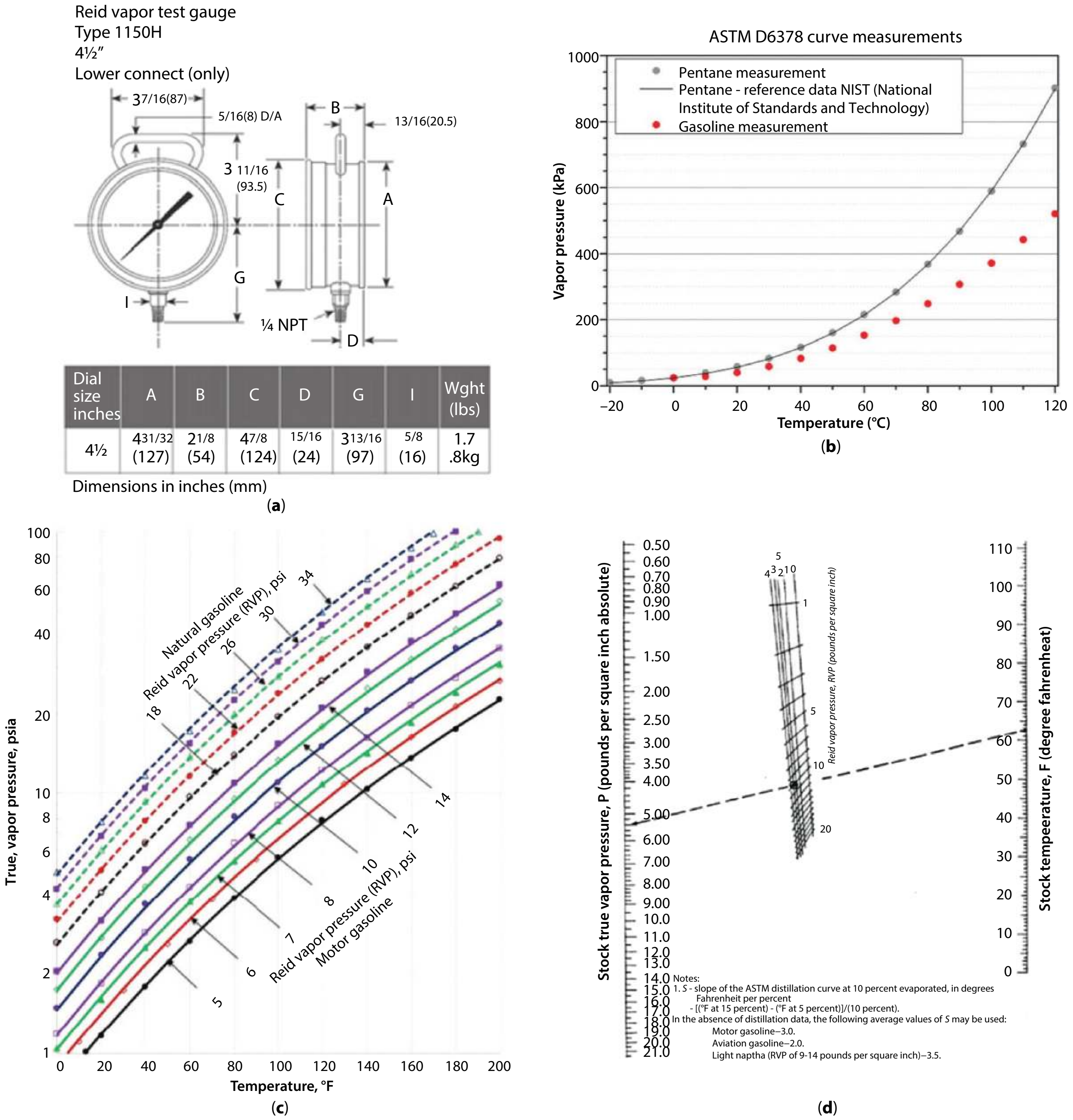
Residual Fuel: Heavy fuel oil made from long, short, or cracked residue plus whatever cutter stock is necessary to meet market specifications.

Residue: The bottoms from a crude oil distilling unit, vacuum flasher, thermal cracker, or visbreaker. See Long Residue and Short Residue.

Residuum: Residue from crude oil after distilling off all but the heaviest components with a boiling range greater than 1000°F (538°C).

Reynolds Number (Re): A dimensionless number, Re, expressing the ratio of inertial to viscous forces in a flowing fluid and can be used to determine the flow regime. For a fluid in a pipe of circular cross-section:

$$\mathrm{Re} = \frac{\rho v d}{\mu}$$



Dial size inches	A	B	C	D	G	I	Wght (lbs)
4½	431/32 (127)	21/8 (54)	47/8 (124)	15/16 (24)	313/16 (97)	5/8 (16)	1.7 .8kg

Dimensions in inches (mm)

(a)

Notes:
1. *S* - slope of the ASTM distillation curve at 10 percent evaporated, in degrees Fahrenheit per percent
- [(°F at 15 percent) - (°F at 5 percent)]/(10 percent).

In the absence of distillation data, the following average values of *S* may be used:
Motor gasoline–3.0.
Aviation gasoline–2.0.
Light naptha (RVP of 9-14 pounds per square inch)–3.5.

Figure 41 (a) Reid vapor test gauge. (b) Vapor pressure vs. temperature. (c) Reid vapor pressure vs. temperature.

where ρ is the density, v is the mean velocity, d is the diameter of a pipe, and μ is the viscosity. Where the value for critical pipes falls below 2000, the flow is laminar flow or streamlined. For a Reynolds number above 4000, the flow is turbulent.

Rich oil: The absorption oil leaving the bottom tray of an absorption column. The rich oil contains the absorbed light components.

Ring Compounds: Hydrocarbon molecules in which the carbon atoms form at least one closed ring such as naphthenes or aromatics. Also called cyclic.

Table 3 RVP blending values.

RVP blending values		Vol% (aromatics)					
	rvp (pure HC)	0	10	20	30	40	50
Ethane	730.0	474.0	474.0	474.0	474.0	474.0	474.0
Propene	226.0	216.0	216.0	216.0	216.0	216.0	216.0
Propane	190.0	173.0	173.0	173.0	173.0	173.0	173.0
Isobutane	72.2	62.0	73.9	85.4	96.6	107.6	118.8
Isobutene	63.4	76.5	78.9	81.3	83.7	86.2	88.9
Butene-1	63.0	76.1	78.4	80.8	82.7	85.1	87.4
n-Butene	51.6	52.9	55.6	58.3	60.9	63.5	66.2
trans-2-Butene	49.8	62.1	64.0	66.0	68.0	70.0	72.0
cis-2-Butene	45.5	58.6	60.5	62.3	64.2	66.1	69.0
Isopentane	20.4	21.9	22.2	22.5	22.9	23.3	23.7
C_5 olefins*	16.5	17.9	18.1	18.4	18.6	18.8	19.0
n-Pentane	15.6	16.9	17.2	17.4	17.8	18.0	18.2

*C_5 olefins in FCC proportion.

Ring Structure: A compound in which some of the carbon atoms are linked with other carbon atoms to form a continuum. Carbon atoms attached to the ring carbon atoms are said to be "side chains."

Riser: **1.** A pipe through which a fluid travels upwards (See Figure 40; a pipe to the reactor). **2.** A steel or flexible pipe that transfer fluids well from the seabed to the surface.

Risk: **1**. Is defined as a measure of economic loss, human injury, or environmental damage in terms of both the incident likelihood and the magnitude of the loss, injury, or damage. **2**. The probability of an event happening times the impact of its occurrence on operations. (Impact is the effect on conditions or people if the hazard is realized (occurs) in practice and potentials are the likelihood that the impact will occur.

Risk Analysis: A decision-making tool that allows an examination of the level and significance of workplace risk for humans, equipment, weather, operations, or other conditions. Determines the probability of the risk occurring, the impact the risk will have, and how to mitigate the risk. *See Hazard Analysis*.

Risk Assessment: The process of identifying and evaluating the technical and non-technical risks associated with a project. It includes the amount or degree of potential danger perceived (by an assessor) when determining a course of action to accomplish a given task. Risk assessment may be qualitative or quantitative.

Risk Matrix: Is the common approach to risk assessment and hazard analysis. Its underlying idea is that acceptability of risk is a product of how likely a thing is to happen and how bad it would be if it happened. This is shown in the following tables.

Category	Definition	Range (failures per year)
Certain	Many times in system lifetime	$>10^{-3}$

(*Continued*)

(***Continued***)

Category	Definition	Range (failures per year)
Probable	Several times in system lifetime	10^{-3} to 10^{-4}
Occasional	Once in system lifetime	10^{-4} to 10^{-5}
Remote	Unlikely in system lifetime	10^{-5} to 10^{-6}
Improbable	Very unlikely to occur	10^{-6} to 10^{-7}
Inconceivable	Cannot believe that it could occur	$<10^{-7}$

Risk matrix categorization of severity of consequences

Category	Definition
Catastrophic	Multiple loss of life
Critical	Loss of a single life
Marginal	Major injuries to one or more persons
Negligible	Minor injuries to one or more persons

Risk matrix

Consequence

Likelihood	Catastrophic	Critical	Marginal	Negligible
Certain	Class I	Class I	Class I	Class II
Possible	Class I	Class I	Class II	Class III
Occasional	Class I	Class II	Class III	Class IV
Remote	Class II	Class III	Class III	Class IV
Improbable	Class III	Class III	Class IV	Class IV
Inconceivable	Class IV	Class IV	Class IV	Class IV

Key:
Class I: Unacceptable
Class II: Undesirable
Class III: Tolerable
Class IV: Acceptable

Road Oil: Any heavy petroleum oil, including residual asphaltic oil used as a dust palliative and surface treatment on roads and highways. It is generally produced in six grades from 0, the most liquid, to 5, the most viscous.

Rule of Thumb: Axioms based on practical experience and/or methods to approximate calculated results using simple formulae.

Runback: The liquid returning to the flash zone of a distillation column.

Safety: A general term denoting an acceptable level of risk of relative freedom from and low probability of harm.

Safeguard: A precautionary measure of stipulation. Usually equipment and/or procedures designed to interfere with incident propagation and/or prevent or reduce incident consequences.

Safety Integrity Level (SIL): 1. Is defined as a relative level of risk reduction provided by a safety function or to specify a target level of risk reduction. SIL is a measure of performance required for a safety instrumented function (SIF). 2. The degree of redundancy and independence from the effects of inherent and operational failures and external conditions that may affect system performance.

The requirements for a given SIL are not consistent among all of the functional safety standards. In the European functional safety standards based on the IEC 61508 standard, four SILs are defined, with SIL 4 the most dependable and SIL 1 the least. An SIL is determined based on a number of quantitative factors in combination with qualitative factors such as development process and safety life cycle management.

The assignment of SIL is an exercise in risk analysis where the risk associated with a specific hazard, which is intended to be protected against by an SIF, is calculated without the beneficial risk reduction effect of the SIF. That "unmitigated" risk is then compared against a tolerable risk target. The difference between the "unmitigated" risk and the tolerable risk, if the "unmitigated" risk is higher than tolerable, must be addressed through the risk reduction of the SIF. This amount of required risk reduction is correlated with the SIL target. In essence, each order of magnitude of risk reduction that is required correlates with an increase in one of the required SIL numbers.

There are several methods used to assign an SIL. These are normally used in combination and may include:

- *Risk matrices*
- *Risk graphs*
- *Layers of Protection Analysis (LOPA)*

Of the methods presented above, LOPA is by far the most commonly used by large industrial facilities.

The assignment may be tested using both pragmatic and controllability approaches, applying guidance on the SIL assignment published by the UK HSE. SIL assignment processes that use the HSE guidance to ratify assignments developed from Risk Matrices have been certified to meet IEC EN 61508.

Safety Instrumented Function (SIF): A safety function with a specific safety integrity level that is necessary to achieve functional safety and that can either be a safety instrumented protection function or a safety instrumented control function.

Salt Content: Crude oil usually contains salts in solution in water that is emulsified with the crude. The salt content is expressed as the solution of sodium chloride (NaCl) equivalent in pounds per thousand barrels (PTB) of crude oil. Typical values range from 1 to 20 PTB. Although there is no simple conversion from PTB to parts per million by weight (ppm), 1 PTB is roughly equivalent to 3 ppm.

Saturated Compounds: Hydrocarbons in which there are no double bonds between carbon atoms. Saturated compounds contain the maximum number of hydrogen atoms that are possible.

Screwed Fittings: These are used to assemble screwed connections and field instruments on pipes. They are:

- Pipe thread fittings
- Instrument or tubing fittings
- Metric fittings

None of these will screw together.

Scrub: The removal of components (gas, liquids, or solids) from methane achieved by surface equipment (scrubbers).

Scrubber: **1**. A reactor that removes various components from produced gas. **2**. Equipment that causes the separation of liquid and gaseous phases in a fluid system. The separation is usually based on the density differences of the two phases and can take place using gravity force, induced centrifugal force, and so on. **3**. A system to reduce noxious substances from a flowing stream of air, usually filled with plates or packing, through which the scrubbing fluid flows countercurrent or cross-current to the path of the contaminated air.

Scrubbing: The purification of a gas or liquid by washing it in a tower.

Secondary Absorber: The second absorber in an FCC gas plant. It is usually the last unit operation in the gas recovery plant and is also known as the sponge absorber.

Selectivity: The difference between the research octane number and the motor octane number of a given gasoline. Alkylate is an excellent low-sensitivity and reformates a high-sensitivity gasoline component. It is an indication of the sensitivity of the fuel to driving conditions (city vs. highway).

Selective Treating: The preferential removal of one acid gas component, leaving at least some of the other acid gas components in the treated stream.

Sensitivity: The difference in the research octane (F-1) and the motor octane (F-2) for a gasoline stream. Since the research octane is always larger, sensitivity is always a positive number.

Separation zone: A section of a distillation column in which a separation between two products occurs. The components that are found in both products are said to be distributed components.

Separator: Usually refers to a drum, in which the residence time is provided for a mixture of the liquid and vapor to separate into liquid and vapor streams. Also called a flash drum. The liquid and vapor leaving the separator are in phase equilibrium.

Severity: The degree of intensity of the operating conditions of a process unit. Severity may be indicated by the clear research octane number of the product (reformer), the percentage disappearance of the feed (catalytic cracking), or operating conditions alone (usually the temperature; the higher the temperature, the greater the severity).

Shale: **1**. A common sedimentary rock with porosity but little matrix permeability. Shales are one of the petroleum source rocks. Shales usually consist of particles finer than sand grade (less than 0.0625 mm) and include both silt and clay grade material. **2**. A very fine-grained sedimentary rock formed by the consolidation and compression of clay, silt, or mud. It has a finely laminated or layered structure. Shale breaks easily into thin parallel layers; a thinly laminated siltstone, mudstone, or claystone. Shale is soft but sufficiently hard-packed (indurated), so as not to disintegrate upon becoming wet. Some shales absorb water and swell considerably, causing problems in well drilling. Most shales are compacted and consequently do not contain commercial quantities of oil and gas. **3**. Rock formed from clay. **4**. Gas reserves found in unusually non-porous rock, requiring special drilling and completion techniques.

Shale Gas: Methane (CH_4) gas stored in shale. May be in the pore space, adsorbed to the mineral or rock surfaces, or as free gas in the natural fractures.

Shale Oil: **1**. Can be either an immature oil phase, often called kerogen or actual oil in the cracks or pores of shale. **2**. The liquid obtained from the destructive distillation of oil shale. Further processing is required to convert it into products similar to petroleum oils.

Shear force: An applied force to a material that acts in a direction that is parallel to a plane rather than perpendicular. A material such as a solid or fluid is deformed by the application of a shear force over a surface known as shear stress. The shear strain is the extent of the deformation defined as the ratio of the deformed distance with length. The shear modulus is the ratio of the shear stress to the shear strain.

Shear rate (γ): The deformation of fluid under the influence of an applied shear force presented as the change in velocity of the fluid perpendicular to flow:

$$\gamma = \frac{dv}{dz}$$

where dv/dz is referred to as the velocity gradient. The S.I. unit is s^{-1}.

Shear stress (τ): The shear force applied to a fluid that is applied over a surface. When the shear stress is proportional to the shear rate, the fluid exhibits Newtonian behavior and the viscosity is constant. The S.I. unit is Nm^{-2}.

$$\tau = \mu \frac{dv}{dz}$$

where μ is the viscosity.

Shell and Tube Heat Exchanger: A device used to transfer heat from one medium to another. It consists of a shell that contains tubes. One medium is contained within the shell and the other within the tubes, and heat is transferred from one to the other across the tubes. There are many designs commonly used, and the simplest is a single phase-type exchanger in which a cold liquid to be heated flows through the tubes from one side of the exchanger to the other. Steam is used as the heating medium and enters as vapor and leaves as condensate from the bottom. A kettle reboiler type is a type of shell and tube heat exchanger in which steam is admitted through the tubes. The choice of hot or cold fluid in the tubes or shell depends on the application and nature of the fluids, such as their susceptibility to fouling (See Figure 42).

Shell Side: The space between the outside of the tubes and the inside of the casing or shell of a shell and tube heat exchanger.

Sherwood Number (Sh): A dimensionless number that represents the relationship between mass diffusivity and molecular diffusivity:

$$Sh = \frac{kL}{D_{AB}}$$

where k is the mass transfer coefficient, L is the characteristic dimension, and D is the diffusivity of the solute A in solvent B. It corresponds to the Nusselt number used in heat transfer.

Shock wave: A pressure wave of very-high-pressure intensity and high temperature that is formed when a fluid flows supersonically or in which a projectile moves supersonically through a stationary fluid. It can be formed by a violent event such as a bomb blast or an explosion. A shock-wave compression is the non-isentropic adiabatic compression in a wave that is traveling above the speed of sound.

Short Residue: Flasher bottoms or residue from the vacuum tower bottoms.

Short-Term Exposure Limit (STEL): The time-weighted average concentration of a substance over a 15-min. period thought not to be injurious to health.

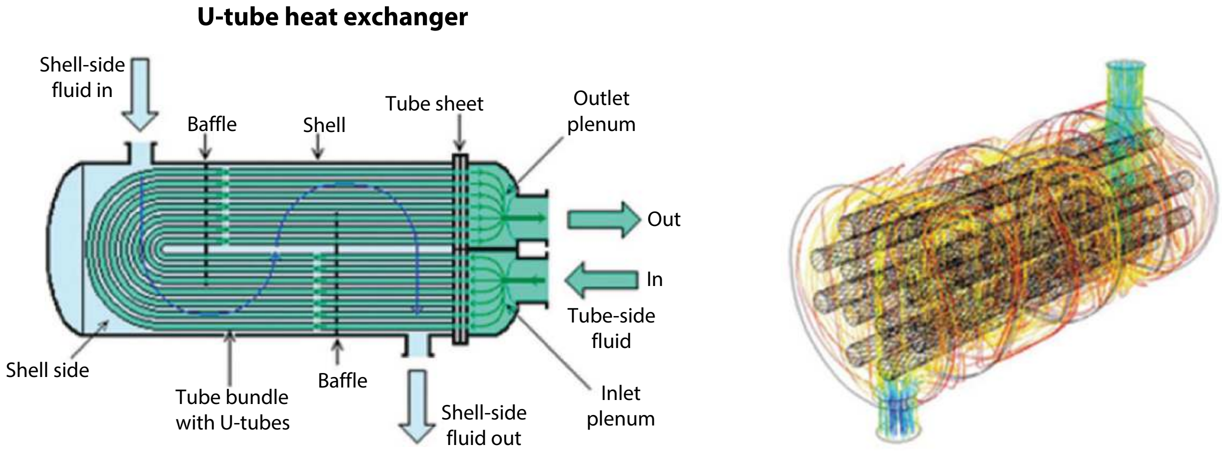


Figure 42 A shell and tube heat exchanger shows the direction of the flow of fluids in the shell and tube sides.

Shutdown: 1. The status of a process that is not currently in operation due to scheduled or unscheduled maintenance, cleaning, or failure. 2. A systematic sequence of action that is needed to stop a process safely.

Side Draw: See Draw

Side Heater (reboiler): A heat input to a distillation column that is located above the bottom tray of the column.

Side Reaction: A chemical reaction that takes place at the same time as a main reaction and produces unwanted products and therefore reduces the yield of the desired product. For example, in the high-temperature cracking reaction of propane (C_3H_8) to produce propylene (C_3H_6), $C_3H_8 \rightarrow C_3H_6 + H_2$, some of the hydrogen can react with the propane to produce methane and ethane as side reactions, $C_3H_8 + H_2 \rightarrow CH_4 + C_2H_6$. The conditions for the reaction must therefore be controlled to reduce this unwanted reaction.

Side Stream: The continuous removal of a liquid or a vapor from a process such as a distillation column that is not the main process flow. For example, drawing off vapor or liquid midway up the column can have an economic advantage in terms of the physical size of the column and the amount of boil-up energy required.

Side Stripper: A small auxiliary column that receives a liquid draw product from the main distillation column for the stripping of light components. Light components are stripped by stripping steam or reboiling and returned to the main column. Liquid products are sometimes stripped inside strippers to raise the flashpoint.

Sieve Plate Column: **1**. A type of distillation column that uses a stack of perforated plates to enhance the distribution and intimate contact between vapor and liquid. The plates allow vapor to pass up and bubble through the liquid on the plates. The rate of flow of vapor is sufficient to prevent the liquid from draining down the sieve plates. Instead, the liquid flows over a weir and down a downcomer to the sieve plate below. **2**. Sieve trays are metal plates with holes; vapor passes straight through the liquid on the plate. The arrangement, number, and size of the holes are designed parameters (See Figure 43).

Simulated distillation (Simdist): A relatively new laboratory technique in which a petroleum stream is separated into fractions with gas-phase chromatography. Carbon disulfide (C_2S) is used as the carrying agent to dissolve the petroleum stream. The component fractions elute from the chromatographic column in a time sequence, related to their boiling temperatures. Temperatures are assigned to the fractions based on the chromatographic separation

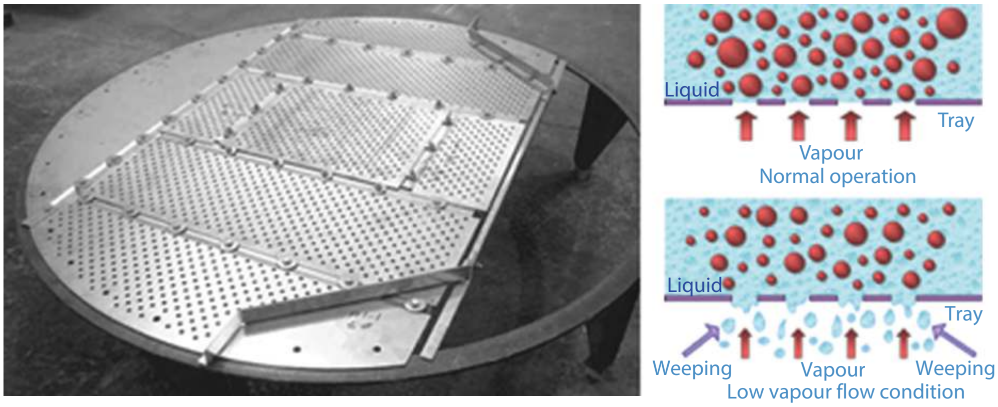


Figure 43 A sieve plate.

of a normal paraffin standard mixture. The simulated distillation approaches a true boiling point distillation and is reported on a mass basis for streams heavier than gasoline. Aromatic compounds elute from the column faster than the paraffin of similar boiling points. Therefore, simulated distillations must be corrected for aromatic content when stocks contain significant quantities of aromatic components.

Slack Wax: Wax produced in the dewaxing of lube oil base stocks. This wax still contains some oil and must be oiled to produce the finished wax product.

Slop Wax: The over flash from a vacuum column. The slop wax is usually withdrawn from the column and combined with the fresh charge to the vacuum furnace.

Slurry: The bottom stream from FCC main fractionators. It is termed slurry because it contains suspended catalyst particles.

Slurry Oil: The oil, from the bottoms of the FCC unit fractionating tower, containing FCC catalyst particles carried over by the vapor from the reactor cyclones. The remainder of the FCC bottoms is the decanted oil (See Figure 40).

Smoke: The gaseous products of the burning of carbonaceous materials made visible by the presence of small particles of carbon; the small particles that are of liquid and solid consistencies are produced as a byproduct of insufficient air supplies to a combustion process.

Smoke Point: **1**. Refers to the height of a smokeless flame of fuel in millimeters beyond which smoking takes place. It reflects the burning quality of kerosene and jet fuels. **2**. A test measuring the burning quality of jet fuels, kerosene, and illuminating oils. It is defined as the height of the flame in millimeters beyond which smoking takes place; ASTM D 1322.

Soaker, Soaking Drum: A soaker is a device that allows cracking time (soaking time) for heated oil in a thermal cracking operation. Furnace coils and/or drums are used for this purpose. Since some coke is deposited in the soaking device, it must be periodically taken offline and cleaned. Furnace coils are much easier to clean than drums.

Soave–Redlich–Kwong (SRK) equation of state: An equation of state widely used to predict the vapor–liquid equilibria of substances. It is a development of the "Redlich–Kwong" equation of state that correlated the vapor pressure of normal fluids:

$$p = \frac{RT}{v-b} - \frac{a\alpha(T)}{V(V+b)}$$

where a and b are constants and obtained from critical point data. It also involves a function that was developed to fit vapor pressure data using reduced temperature, T_r:

$$\alpha = \left[1 + \left(0.480 + 1.574\omega - 0.176\omega^2\right)\left(1 - T_r^{0.5}\right)\right]^2$$

where ω is the acentric factor.

Solvent Extraction: A separation process based on selective solubility, where a liquid solvent is introduced at the top of a column. As it passes the feed, which enters near the bottom as a vapor, it selectively dissolves a target constituent. The solvent is then removed via the bottom of the column and put through an easy solvent/extract fractionation. From the top of the column comes a raffinate stream, the feed stripped out of the extract. Butadienes and aromatics are some products recovered by solvent extraction (Figure 44).

Sour Crude Oils: Crudes that contain sulfur in amounts greater than 0.5–1.0 wt % or that contain 0.05 ft.^{3} or more hydrogen sulfide (H_2S) per 100 gal. Such oils are dangerously toxic. Even 0.05 ft.^{3} per 100 gal can be present before severe corrosion tends to occur. Arabian crudes are high-sulfur crudes that are not always considered sour because they do not contain highly active sulfur compounds. The original definition was for any crude oil that smelled like rotten eggs.

Sour Gas: **1**. A light gas stream that contains acid gases, in particular sulfur compounds, ammonia compounds, and carbon dioxide. **2**. Gas rich in hydrogen sulfide (H_2S). **3**. Natural gas that contains a significant amount of hydrogen sulfide (usually greater than 16 ppm) and possibly other objectionable sulfur compounds (mercaptans, carbonyl sulfide). Also called "acid gas." **4**. Natural or associated gas with high sulfur content. **5**. Natural gas containing chemical

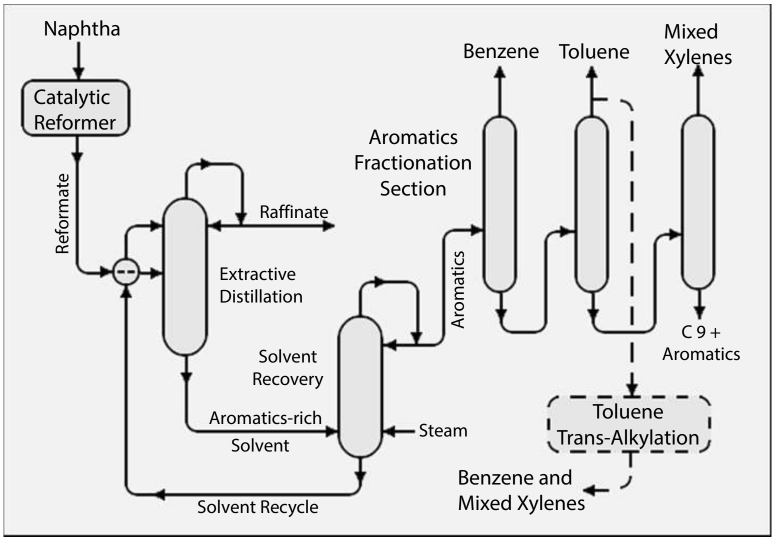


Figure 44 Solvent extraction involving aromatics (BTX).

impurities, a notable hydrogen sulfide (H_2S), or other sulfur compounds that make it extremely harmful to breathe even small amounts; a gas with a disagreeable odor resembling that of rotten eggs. **6**. A gas containing sulfur-bearing compounds such as hydrogen sulfide and mercaptans and that is usually corrosive. **7**. Raw natural gas to be processed, that is, gas received at the liquefaction plant before being subjected to any pretreatment.

Space Velocity: A unit generally used for expressing the relationship between the feed rate and reactor volume in a flow process. It is defined as the volume or weight of feed per unit time per unit volume of the reactor or per unit weight of the catalyst. Space velocity is normally expressed on a volume basis (LHSV: liquid hourly space velocity) or a weight basis (WHSV: weight hourly space velocity). The LHSV and WHSV are determined as follows:

$$\text{LHSV} = \frac{\text{total volumetric feed flow rate to the reactor}}{\text{total catalyst volume}} [=]\, h^{-1}$$

$$\text{WHSV} = \frac{\text{total mass feed flow rate to the reactor}}{\text{total catalyst weight}} [=]\, h^{-1}$$

LHSV and WHSV are related by the equation:

$$\text{WHSV} = \frac{\rho_{\text{oil}}}{\rho_{\text{cat}}} \text{LHSV}$$

where ρ_{oil} and ρ_{cat} are the densities of the hydrocarbon feed and the catalyst, respectively.

Specific gravity: By definition, is the ratio of gas density (at the temperature and pressure of the gas) to the density of dry air (at the air temperature and pressure).

Spent Catalyst: A catalyst that has been through a reaction and is no longer as active because of substances or other contaminants deposited on it (in the case of solid) or mixed with it (in the case of liquid).

Spillback: A spillback allows fluid to recycle from the discharge back to the suction of a machine. It is one way to stop a centrifugal compressor from surging.

Splitter: A distillation column that separates a feed into light and heavy products.

Sponge Absorber: See Secondary Absorber.

Sponge Oil: The liquid used in an absorption plant to soak up the constituent to be extracted.

Stability: Is the ability of a catalyst to maintain its activity and selectivity over a reasonable period. A catalyst with good stability has a long cycle life between regeneration in a commercial unit.

Stabilization: A process for separating the gaseous and more volatile liquid hydrocarbons from crude petroleum or gasoline and leaving a stable (less volatile) liquid so that it can be handled or stored with less change in composition.

Stabilizer: A distillation column that removes light components from a liquid product. This terminology is often used to describe debutanizer columns that remove C_4 hydrocarbons from gasoline to control the vapor pressure.

Standard cubic feet (scf): The volume of gas expressed as standard cubic feet. Standard conditions in petroleum and natural gas usage refer to a pressure base of 14.696 psia (101.5 kPa) and a temperature base of 60°F (15°C).

Static head: The potential energy of a liquid expressed in the head form:

$$h = \frac{p}{\rho g}$$

where p is the pressure, ρ is the density, and g is the acceleration due to gravity. It is used directly in the Bernoulli equation for which the other two head forms are velocity head and pressure head.

Steady State: Describes a process in which the mass and energy flowing both into and out the process are in perfect balance.

Steam: The gaseous form of water formed when water boils. At atmospheric pressure, steam is produced at 212°F (100°C) by boiling water. It is widely used in the chemical and process industries as a utility for heating processes such as a kettle-type reboiler for distillation columns. It is also used in power generation when steam is produced or raised from a thermal process and expanded through turbines. Other uses of steam at destroying harmful pathogens and is a harmless substance once cooled. Wet steam is water vapor that contains water droplets. With further heating, the water evaporates. The ***dryness fraction*** of steam is the ratio of the amount of water in steam to the total amount of water vapor. Superheated steam is produced by heating the steam above the boiling point of water. The thermodynamic properties of steam are presented in the published literature as steam tables.

Steam (purchased): Steam, purchased for use by a refinery that was not generated from within the refinery complex.

Steam Cracking: **1**. The high-temperature reduction in length or cracking of long-chain hydrocarbons in the presence of steam to produce shorter-chain products such as ethylene (C_2H_4), propylene (C_3H_6), and other small-chain alkenes (C_nH_{2n}). **2**. The same as catalytic cracking, but specifically refers to the steam injected with the catalyst and feed to give the mixture lift up the riser.

Steam Distillation: The separation of immiscible organic liquids by distillation using steam. It involves the injection of live steam into the bottom of the distillation column and into the heated mixture for separation. The steam reduces the partial pressure of the mixture and the temperature required for vaporization. When distilled, the components operate independently of one another, with each being in equilibrium with its own vapor. Steam distillation is used in the primary separation of crude distillation in a fractionating column.

Steam Injection: The use of live steam fed directly into a process to provide water and heat and to enhance either reaction or extraction. It is commonly used as an enhanced oil recovery method to recover oil from depleted reservoirs or from oil sands in which viscous heavy oil is recovered using steam injection to reduce the viscosity of the oil and aid transport and recovery. Steam is also directly used in the separation of crude oil and fed to the bottom of the fractionating/distillation column. This is the primary separation of crude oil into fractions that have different boiling points. Steam cracking uses steam for the thermal cracking and reforming of hydrocarbons.

Steam jet ejector: A type of fixed operating pump that uses high-pressure steam passed through a constriction to create a low pressure due to the venture effect and to which the equipment to be evacuated is connected such as a distillation column condenser. In spite of requiring high-pressure steam, the device has no moving parts and therefore has low maintenance costs. It can handle corrosive vapors.

Steam Methane Reformer: A primary source of hydrogen in a refinery, this operating unit converts methane (CH_4) and steam (H_2O) to hydrogen (H_2) with the byproducts of carbon monoxide (CO) and carbon dioxide (CO_2).

Steam Point: The temperature that corresponds to the maximum vapor pressure of water at standard atmospheric pressure (1.01325 bar). This corresponds to a temperature of 100°C.

Steam Reforming: The conversion of methane (CH_4) from natural gas into hydrogen (H_2). It is used in the production of ammonia (NH_3) in which the methane is first produced from desulfurized and scrubbed natural gas, mixed with steam, and passed over a nickel catalyst packed in tubes at a high temperature of around 1652°F (990°C)

$$CH_4 + H_2O \rightarrow CO + 3H_2$$

$$CH_4 + 2H_2O \rightarrow CO_2 + 4H_2$$

The reactions are endothermic (i.e., absorbing heat).

Steam Tables: Published tables that present thermodynamic data for enthalpy, entropy, and the specific volume of steam at various temperatures and pressures. Steam is a commonly encountered material in chemical processes, and its properties have been extensively tabulated.

Steam Tracing: An internal pipe or tube used in process vessels and pipelines carrying steam to provide adequate heating to a fluid to keep it at a controlled temperature. The amount of steam or heat supplied is sufficient to overcome losses. Steam tracing is typically used in pipelines carrying molten bitumen and other fluids prone to solidification on cooling, to ensure that they remain in a liquid state.

Steam Trap: A device used to automatically drain and remove condensate from steam lines to protect the steam main from condensate buildup. Various types of steam traps are used and generally consist of a valve that can be operated by a float, spring, or bellows arrangement. The discharge of the hot condensate may be either to the environment or into a collection pipe and returned to the boiler for reuse (See Figure 45).

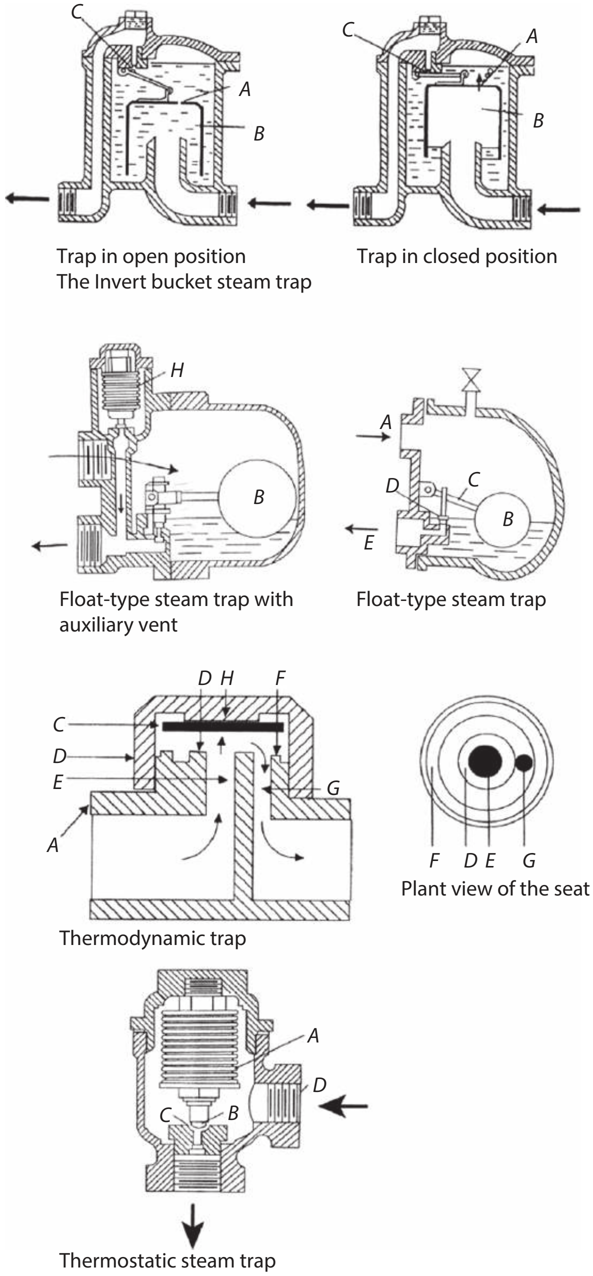
Still Gas (refinery gas): Any form or mixture of gases produced in refineries by distillation, cracking, reforming, and other processes. The principal constituents are methane (CH_4), ethane (C_2H_6), ethylene (C_2H_4), normal butane (nC_4H_{10}), butylenes (C_4H_8), propane (C_3H_8), propylene (C_3H_6), and so on. Still gas is used as a refinery fuel and a petrochemical feedstock. The conversion factor is 6 million Btu per fuel oil-equivalent barrel.

Stoichiometric: Applied to reactors in which the reactants and products are defined in terms of the molar quantities reacting. For example, in the reaction: $3H_2 + 2N_2 \Leftrightarrow 2NH_3$, the stoichiometric coefficients are -3.0, -2.0, and 2.0 for the H_2, N_2, and NH_3, respectively.

Straight-Run Distillate or Natural Gasoline: **1**. A fraction obtained on the simple distillation of crude oil without cracking. Its octane number is usually low and thus requires upgrading by catalytic reforming. **2**. A product that has been distilled from crude oil but has not been through a process in which the composition has been chemically altered. **3**. Gasoline produced by the primary distillation of crude oil. It contains no cracked, polymerized, alkylated, reformed, or visbroken stock.

Stress Relief: Coded vessels typically have a metal stamp attached that states "Do not weld, stress relieved." That means the vessel has been post-weld heat-treated to remove stresses in the vessel wall created by welding during fabrication.

Stripping: The removal (by steam-induced vaporization or flash evaporation) of the more volatile components from a cut or fraction.



Trap in open position
The Invert bucket steam trap

Trap in closed position

Float-type steam trap with auxiliary vent

Float-type steam trap

Thermodynamic trap

Plant view of the seat

Thermostatic steam trap

Figure 45 Steam traps.

Stripper Column: A loose designation applied to a distillation column in which light components are stripped from a heavier liquid product.

Stripping Steam: Steam that is injected into the bottom of a side-stripping column or used to strip oil from the catalyst in an FCC operation.

Stripping Zone: The section of the column in which light components are stripped from a heavier liquid product. In conventional distillation columns, this is the portion of the column from the reboiler to the feed tray.

Sulfolane $(CH_2)_4SO_2$: A chemical used as a solvent in extraction and extractive distillation processes (See Figure 44).

Sulfur: A yellowish non-metallic element, sometimes known as "brimstone." It is present at various levels of concentration in many fossil fuels whose combustion releases sulfur compounds that are considered harmful to the environment. Some of the most commonly used fossil fuels are categorized according to their sulfur content, with lower-sulfur fuels usually selling at a higher price. Note: No. 2 distillate fuel is currently reported as having either a 0.05% or lower sulfur level for on-highway vehicle use or a greater than 0.05% sulfur level for off-highway use, home heating oil, and commercial and industrial uses. This also includes Ultra-Low Sulfur Diesel (<15 ppm sulfur). Residual fuel, regardless of use, is classified as having either no more than 1% sulfur or greater than 1% sulfur. Coal is also classified as being low sulfur at a concentration of 1% or less or high sulfur at concentrations greater than 1%.

Sulfuric Acid Treating: A refining process in which unfinished petroleum products such as gasoline, kerosene, and lubricating oil stocks are treated with sulfuric acid to improve their color, odor, and other characteristics.

Sulfurization: Combining sulfur compounds with petroleum lubricants.

Sulfur Content: Is expressed as a percentage of sulfur by weight and varies from less than 0.1% to greater than 5%. Crude oils with less than 1 wt% sulfur are called low-sulfur content or sweet crude, and those with more than 1 wt% sulfur are called high-sulfur content or sour crude. The sulfur-containing constituents of the crude oil include simple mercaptans (also known as thiols), sulfides, and polycyclic sulfides. Mercaptan sulfur is simply an alkyl chain (R-) with an SH group attached to it at the end. The simplest form of R – SH is methyl mercaptan, CH_3SH.

Surface Area: The total area that a solid catalyst exposes to the feeds in a reaction. The surface area is enhanced in some catalysts like zeolites by extensive microscopic pores.

Supply: The components of petroleum supply are field production, refinery production, imports, and net receipts when calculated on a PADD basis.

Surge: This is a terrifying sound that centrifugal compressors make when they malfunction either due to low flow or excessive discharge pressure or low-molecular-weight gas.

Sweet Crude: **1**. Crude oil containing very little sulfur and having a good odor. **2**. Crude petroleum containing little sulfur with no offensive odor. **3**. Gets its name due to a pleasant and "sweet" smell. Sweet crude has a sulfur content of less than 1%. It is more valuable than sour crude because it costs less to process the crude into finished products. **4**. Oil containing little or no sulfur, especially little or no hydrogen sulfide. The original definition was for any crude oil that did not have a bad odor.

Sweetening: **1**. The removal or conversion to innocuous substances of sulfur compounds in a petroleum product by any of a number of processes (doctor treating, caustic and water washing, etc.). **2**. Processes that either remove obnoxious sulfur compounds (primarily hydrogen sulfide, mercaptans, and thiophenes) from petroleum fractions or streams or convert them, as in the case of mercaptans, to odorless disulfides to improve odor, color, and oxidation stability.

Sweet Gas: **1**. Gas sweetened. Gas processed in the acid gas removal unit that no longer contains gaseous pollutants. **2**. Natural gas that contains small amounts of hydrogen sulfide (and other sulfur compounds) and carbon dioxide that can be transported or used without purifying with no deleterious effect on piping and equipment. **3**. A gas stream from which the sulfur compounds have been removed. **4**. A gas containing no corrosive components such as hydrogen sulfide and mercaptans.

Symbols of Chemical Apparatus and Equipment: Listed below are some symbols of chemical apparatus and equipment normally used in a P & ID, according to ISO 10628 and ISO 14617 (See Figure 46).

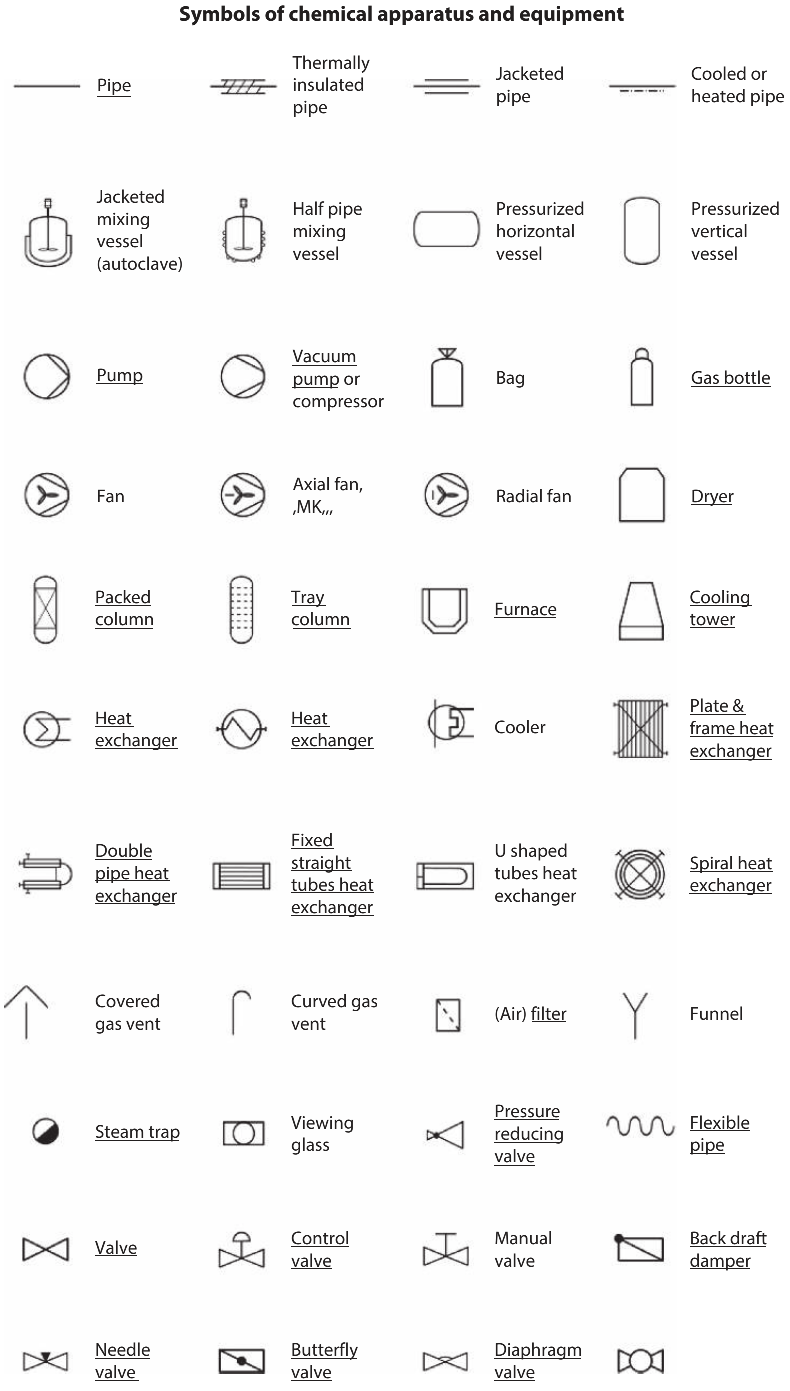


Figure 46 Symbols of chemical apparatus and equipment.

Synthetic Crude: A wide-boiling-range product of catalytic cracking, coking, hydrocracking, or some other chemical structure change operation.

Synthesis Gas: The product of a reforming operation in which hydrocarbons, usually methane and water, are chemically rearranged to produce carbon monoxide, carbon dioxide, and hydrogen. The composition of the product stream can be varied to fit the needs of hydrogen and carbon monoxide at refineries or a chemical plant. Also known as syn gas.

Tail Ends: Small amounts of hydrocarbon in a cut that vaporizes slightly outside the effective initial boiling point and the effective end point.

Tail Gas: Light gases C_1 to C_3 and H_2 produced as byproducts of refinery processing.

TAN: Total acid number.

Tank Farm: An installation used by gathering trunk pipeline companies, crude oil producers, and terminal operators (except refineries) to store crude oil.

Tanker and Barge: Vessels that transport crude oil or petroleum products. Data are reported for movements between PAD Districts; from a PAD District to the Panama Canal; or from the Panama Canal to a PAD District.

Tar: Complex, large molecules of predominantly carbon with some hydrogen and miscellaneous other elements that generally deteriorate the quality of processes and the apparatus.

TBP Distillation: See Fifteen-Five Distillation.

Tertiary Amyl Methyl Ether $(CH_3)_2(C_2H_5)COCH_3$ (TAME): A high-octane oxygenate blending stock produced by reacting isoamylene (isopentylene) produced in FCC processes with methanol. Used to enhance the octane of a motor gasoline pool.

Tertiary Butyl Alcohol – $(CH_3)_3COH$ (TBA): An alcohol primarily used as a chemical feedstock, solvent, or feedstock for isobutylene production for MTBE; produced as a co-product of propylene oxide production or by direct hydration of isobutylene.

Tetra Ethyl Lead (TEL): A compound added to gasoline to increase the octane. TEL has been superseded by other octane enhancers and is no longer used by refiners for motor gasoline.

Test Run: A time period during which the operating data and stream samples are collected for a process. During test runs, the operation of the processing unit is held as steady as possible. For good test runs, the average conditions and stream flows approximate a steady-state operation.

The Diesel Engine: The diesel engine is a reciprocating internal combustion engine. It is different from the petrol engine in that the air intake in the engine cylinder is unthrottled and not premixed with the fuel. Here, the ignition takes place spontaneously without the help of a spark plug. The air taken into the cylinder at atmospheric pressure is compressed to a volume ratio somewhere near to 1:16. At the end of the compression, fuel is injected into the cylinder. The quantity injected depends on the power of the engine, and the heat of compression heats the mass of the air compressed.

The Onion Model: The onion model depicts hazards, barriers, and recovery measures. It reflects the layers of protection and shows how the various measures fit together when viewed from the perspective of the hazard. The first layer is the basic containment of the feedstock, processes, and products (See Figure 47).

The Saybolt Universal Viscometer: Measures the time in seconds that would be required for the total flow of 60 cc of oil from a container tube at a given constant temperature through a calibrated orifice placed at the bottom of the tube. Lubricant viscosities are usually measured in Saybolt Universal seconds at 100°F (37.8°C), 130°F (54.4°C), or 210°F (98.9°C).

For example, the symbol SSU 100 represents the time in seconds that a fluid at 100°F (37.8°C) will take to flow through a given orifice of the Saybolt viscometer.

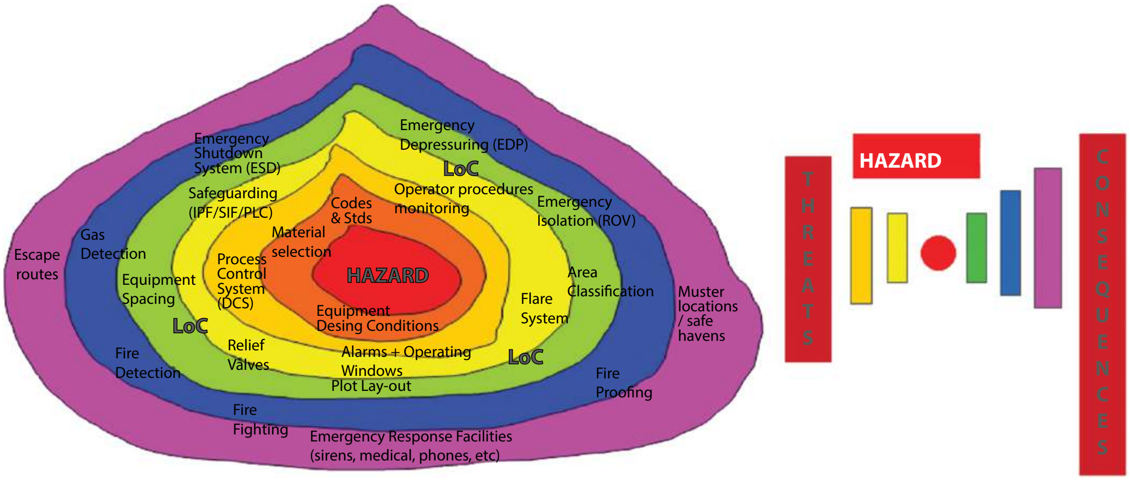


Figure 47 The onion model (LOC = loss of containment).

Kinematic viscosity can be converted to Saybolt viscosity SSU by the formula:

$$\text{Kinematic viscosity}, \nu = \frac{\mu}{\rho} = 0.219t - \frac{149.7}{t}$$

where
μ = viscosity of fluid in centipoises, cP
ρ = density of fluid, g/cc
t = Saybolt Universal viscosity, sec

Theoretical Plate: **1.** A theoretical contacting unit useful in distillation, absorption, and extraction calculations. The vapor and liquid leaving any such unit are required to be in equilibrium under the conditions of temperature and pressure that apply. An actual fractionator tray or plate is generally less effective than a theoretical plate. The ratio of a number of theoretical plates required to perform a given distillation separation to the number of actual plates used, given the overall tray efficiency of the fractionator. **2**. Refers to a vapor/liquid contact device (e.g., distillation column) in which the liquid and vapor leaving the device are in perfect vapor/liquid-phase equilibrium. There are also perfect energy and mass balances for a theoretical tray.

Thermal Cracking: **1**. A refining process in which heat and pressure are used to break down, rearrange, or combine hydrocarbon molecules. Thermal cracking includes gas oil, visbreaking, fluid coking, delayed coking, and other thermal cracking processes (e.g., Flexicoking). **2**. The first cracking process in which the oil was cracked by heating only. Thermal cracking produces a lower-octane gasoline than catalytic processes.

Thermal Cracked Distillate: Is formed when a distillate heavier than gasoline is heated under pressure in order to break the heavy molecules into lighter ones that boil in the gasoline range. This is superseded by catalytic cracking, which gives a better distillate.

Thermal Conductivity: The ability of a material to let heat pass. Metals, water, and materials that are good conductors of electricity have a high thermal conductivity. Air, rubber, and materials that are bad conductors of electricity have a low thermal conductivity. High-viscosity hydrocarbons are bad conductors of heat.

Thermal Expansion: Railroad tracks grow longer in the heat of the sun. The hot tubes in an exchanger grow more than the cold shell. Hence, we have a floating head in the tube bundle to accommodate differential rates of thermal expansion between the tube bundle and the shell.

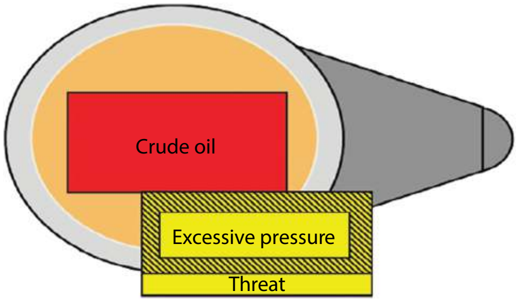


Figure 48 A threat.

Threat: A threat is something that can cause the release of a hazard and lead to a top event (See Figure 48).

Three Phase: A mixture consisting of one vapor in equilibrium with two mutually insoluble liquid phases.

Threshold Limit Value: The amount of a contaminant to which a person can have repeated exposure for an eight-hour day without adverse effects.

Toluene ($C_6H_5CH_3$): **1**. A colorless liquid of the aromatic group of petroleum hydrocarbons, made by the catalytic reforming of petroleum naphthas containing methyl cyclohexane ($CH_3C_6H_{11}$). A high-octane gasoline blending agent, solvent, and chemical intermediate, base for TNT. **2**. One of the aromatic compounds used as a chemical feedstock most notoriously for the manufacture of TNT, trinitrotoluene.

Top Event: A top event is the "release" of the hazard, i.e., the first consequence, typically a loss of containment, a loss of control, or an exposure to something that may cause harm, such as the release of hydrocarbons, toxic substances, or energy (See Figure 49).

Top Product: For columns with condensers, the liquid and/or vapor streams from the reflux drum that exit the process.

Topping: Removal by distillation of the light products and transportation fuel products from crude oil, leaving in the still bottoms all of the components with boiling ranges greater than diesel fuel.

Topped Crude Oil: **1.** Crude that has been run through a distilling unit to remove the gas oil and lighter streams. The long residue is sold as residual fuel. **2**. The bottom product from a crude distillation column.

Toxic Compounds: NO_x, VOCs, and SO_x are toxic compounds such as formaldehyde, oxides of nitrogen, volatile organic compounds such as pentene, and oxides of sulfur.

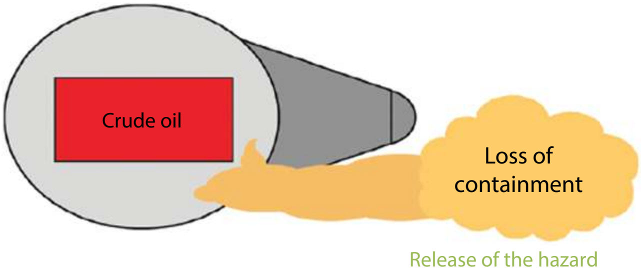


Figure 49 Top event.

Tray: A liquid/vapor contact device in a distillation column (Figure 43).

Tray Efficiency: See Overall Tray Efficiency.

Treat Gas: Light gases, usually high in hydrogen content, which are required for refinery hydrotreating processes such as hydrodesulfurization. The treat gas for hydrodesulfurization is usually the tail gas from catalytic reforming or the product from a hydrogen unit.

Trip: **1**. The fast shutdown of an item of chemical plant or process equipment such as a pump or compressor. The shutdown is the result of a process condition being exceeded such as an abnormal flow, pressure, temperature or concentration, etc. **2**. This a safety device that automatically shuts down a piece of equipment. It is a fail-safe mechanism often activated by unlatching a spring-operated valve, which then closes.

Troubleshooting: A form of problem-solving methodology used to identify, solve, and eliminate problems within a process that has failed or has the potential to fail. It is a logical and systematic search for the source or cause of the problem and solutions presented to ensure that the process is restored back to its full operability. Troubleshooting is applied once a problem has occurred and the process stops functioning. It can take the form of a systematic checklist and requires critical thinking. Computer techniques are employed for more complex systems where a sequential approach is either too lengthy or not practical or where the interaction between the elements in the system is not obvious.

True Boiling Point Distillation (TBP): 1. Of a crude oil or petroleum fractions results from using the US Bureau of Mines Hempel method and the ASTM D-285 test procedures. Neither of these methods specifies the number of theoretical stages or the molar reflux ratio used in the distillation. Consequently, there is a trend toward applying a 15:5 distillation according to ASTM D2892, instead of the TBP. The 15:5 distillation uses 15 theoretical stages and a molar reflux ratio of 5. **2**. A laboratory test in which petroleum oil is distilled in a column having at least 15 theoretical trays and a reflux ratio of 5.0. The distillate is continually removed and further analyzed. The separation corresponds somewhat to a component-by-component separation and is a good measure of the true composition for the sample being distilled. As the temperatures in the still increase, the pressure of the still is lowered to suppress thermal cracking of the sample.

The minimum pressure for most TBP stills is about 38 mm Hg. This allows the distillation of petroleum components boiling up to about 900–950°F (483–510°C) at a pressure of 1 atm. Surprisingly, the TBP test has never been standardized and several different apparatuses are used for the test.

A key result from a distillation test is the boiling point curve, i.e., the boiling point of the oil fraction versus the fraction of oil vaporized. The initial boiling point (IBP) is defined as the temperature at which the first drop of liquid leaves the condenser tube of the distillation apparatus. The final boiling point or the end point (EP) is the highest temperature recorded in the test.

Additionally, oil fractions tend to decompose or crack at a temperature of approximately 650°F (344°C) at 1 atm. Thus, the pressure of TBP distillation is gradually reduced to as low as 40 mmHg, as this temperature is approached to avoid cracking the sample and distorting measurements of true components in the oil.

Tube Bundle: Pipes in a shell-and-tube heat exchanger that are packed into an arrangement to ensure effective heat transfer from the outer surface and good transport for fluids through the tubes. The tubes in the tube bundle are spaced and typically set with a rectangular or triangular pitch and held and sealed with a tube plate. Baffles also provide rigidity and encourage a turbulent flow of fluids through the shell side. The tubes can be a straight single-pass or hairpin double-pass arrangement. The tube bundle can be removed from the shell for periodic cleaning. Lugs are welded to the baffles for lifting purposes (Figures 19 and 42).

Tube Size: Tubing sizes are entirely different from pipe sizes. Tubing is often used in heat exchangers and fired equipment.

Tube Still: See Pipe Still.

Turbine: **1**. A machine used to generate electricity by the expansion of a gas or vapor at high pressure through a set of blades attached to a rotor. The blades rotate as the result of the expansion and conversion of energy. Gas turbines and steam turbines are commonly used to generate electricity. A nozzle is used to direct the high-speed gas or steam over a row of turbine blades. The fluid pushes the blades forward, causing them to rotate due to the change in momentum. A row of stationary blades within the turbine redirects the fluid in the correct direction again before it passes through another set of nozzles and expands to a lower pressure. A steam turbine may have several pressure sections and operate at high-pressure, medium-pressure, and, as the steam expands, a low-pressure section, all linked to the same shaft. The steam in the medium-pressure section may be returned to a boiler and reheated before doing further work to prevent the formation of water in the turbine. **2**. A turbine uses steam pressure or burning gas to drive pumps and compressors at variable speeds. Motor drives are usually fixed-speed machines. Variable speed is an energy-efficient way to control flows by eliminating the downstream control valve.

Turbulent Flow: A fluid flow regime characterized by the fluctuating motion and erratic paths of particles. In pipes of circular cross-section, this occurs at a Reynolds number in excess of 4000. Turbulent flow occurs when inertial forces predominate, resulting in a macroscopic mixing of the fluid.

Turnaround: A planned complete shutdown of an entire process or section of a refinery, or of an entire refinery, to perform major maintenance, overhaul, and repair operations and to inspect, test, and replace process materials and equipment.

Two Phase: A mixture consisting of one vapor in equilibrium with one homogeneous liquid phase.

ULSD: Ultra-low-sulfur diesel. Diesel fuel with <15 ppm sulfur.

Unaccounted for Crude Oil: Represents the arithmetic difference between the calculated supply and the calculated disposition of crude oil. The calculated supply is the sum of crude oil production plus imports minus changes in crude oil stocks. The calculated disposition of crude oil is the sum of crude oil input to refineries, crude oil exports, crude oil burned as fuel, and crude oil losses.

Undistributed Component: A component in a distillation column separation zone that is totally recovered in only one of the products.

Unfinished Oils: All oils requiring further processing, except those requiring only mechanical blending. Unfinished oils are produced by partial refining of crude oil and include naphthas and lighter oils, kerosene and light gas oils, heavy gas oils, and residuum.

Unfractionated Streams: Mixtures of unsegregated natural gas liquid components, excluding those plant condensates. This product is extracted from natural gas.

Unsaturated Compounds: Hydrocarbon compounds in which some of the carbon atoms have multiple bonds with other carbon atoms because of the lack of hydrogen atoms to satisfy the carbon atoms' valences.

Upper Explosive Limit (UEL): The maximum concentration of vapor in air above which the propagation of flame will not occur in the presence of an ignition source. Also referred to as the upper flammable limit or upper explosion limit.

Utilities: Most plants have some of the following utility systems connected to process units.

- Natural gas
- Nitrogen
- Plant air
- Instrument air
- Steam of various pressures
- Cooling water
- Service water
- Boiler feed water
- Fire water
- Fuel gas
- City water

Your company safety policy does not permit you to cross-connect these systems. Connecting natural gas to plant air killed 17 workers at a Louisiana refinery.

Vacuum Distillation: **1**. Distillation under reduced pressure (less than atmospheric), which lowers the boiling temperature of the liquid being distilled. This technique prevents the cracking or decomposition of the charge stock, which occurs above 1000°F (538°C). **2**. A distillation column that operates at sub-atmospheric pressure. Vacuum distillation permits the further distillation of heavy feedstocks at reduced temperatures that minimize cracking reactions.

Vacuum Gas Oil (VGO): A side stream from the vacuum distillation tower.

Vacuum Residuum: The heaviest product from a vacuum distillation column.

Valves: A valve is a device that regulates, directs, or controls the flow of a fluid (gases, liquids, fluidized solids, or slurries) by opening, closing, or partially obstructing various passageways.

Diaphragm Valve: A type of device in which a flexible membrane is used to restrict the rate of flow. The membrane is usually made from a flexible natural or synthetic rubber. Diaphragm valves are typically used for fluids that contain suspended solids (See Figure 50).

Gate Valve: This valve closes by sliding a plate or gate down between two grooves. Used to isolate different portions of the process equipment not used to control flow. The valve closes clockwise and takes about a dozen turns to close. Ninety percent of the valves used in process plants are gate valves (See Figure 51).

Globe Valve: A device that regulates the flow of a fluid in a pipe and consists of a flat disc that sits on a fixed ring seat. The disc is movable and allows flow through the valve (See Figure 52).

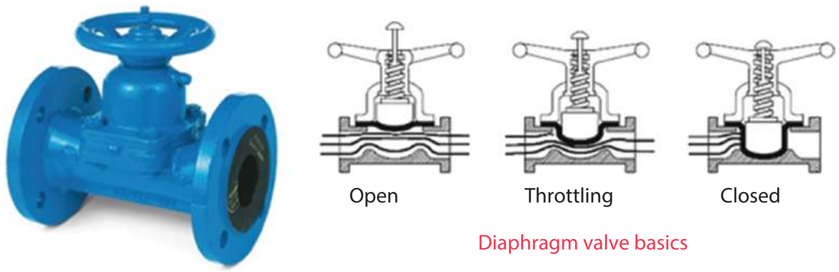


Figure 50 A diaphragm valve.

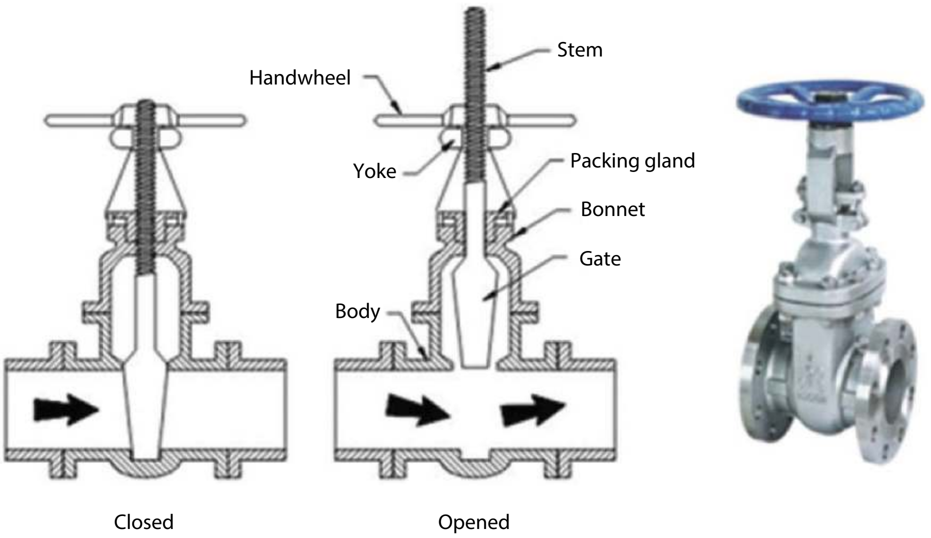


Figure 51 A gate valve.

Plug Valve: This valve goes from 100% open to shut by turning a valve 90°. The natural gas supply to your house is shut off with a plug valve (See Figure 53).

Control Valve: This valve is used to alter flows remotely. Normally, it is moved by air pressure. A gate valve is sometimes used to control flows locally, but this wears out the valve and is best avoided (See Figure 54).

Relief Valves: These valves open to relieve excess pressure to protect a vessel from failure. Also called safety or pop valves (See Figure 55).

Valve Trays: **1**. Fractionator trays that have perforations covered by discs that operate as valves and allow the upward passage of vapor. **2**. In valve trays, perforations are covered by lift-able caps. Vapor flows lift the caps, thus

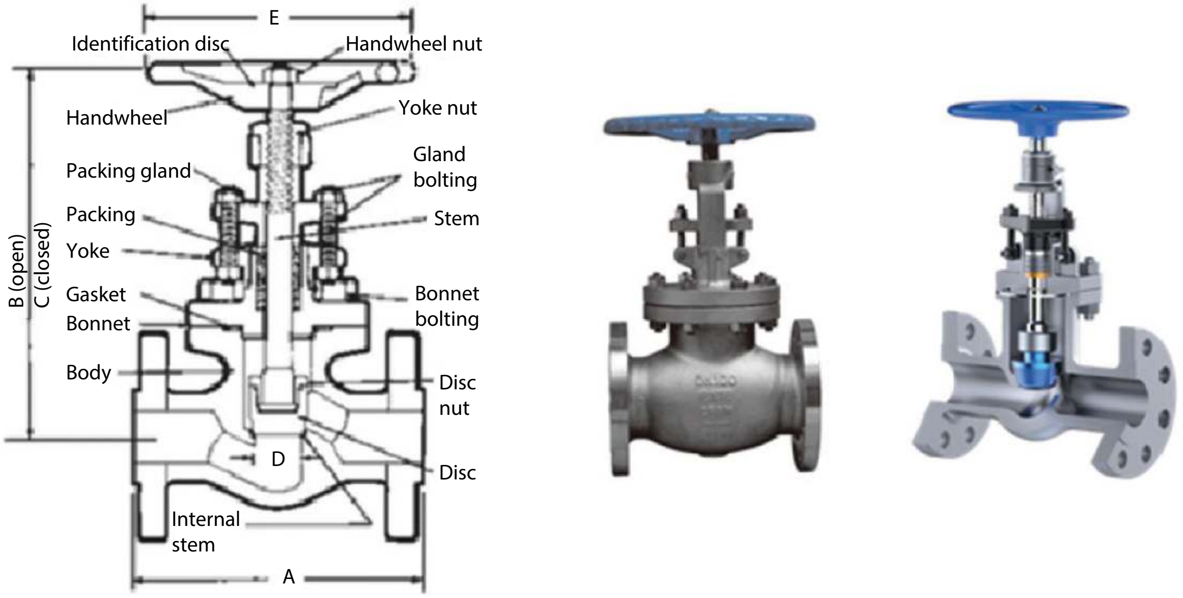


Figure 52 A globe valve away section of a globe valve.

Figure 53 Plug valves. Cutaway section of a plug valve.

self-creating a flow area for the passage of vapor. The lifting caps direct the vapor to flow horizontally into the liquid, thus providing better mixing than is possible in sieve trays (See Figure 56).

Vapor: The gaseous phase of a substance that is a liquid at normal temperature and pressure.

Vapor/Liquid Ratio (V/L): The vapor/liquid ratio (V/L) is the ratio of the volume of vapor formed at atmospheric pressure to the volume of gasoline in a standard test apparatus. The vapor-lock tendency of the gasoline sample can be measured more reliably in terms of its V/L ratio than in terms of its vapor pressure. The V/L ratio also increases with a rise in temperature.

Vapor Pressure: **1**. As a liquid is heated, the molecules in the liquid try to escape into the vapor phase. The hotter the liquid, the harder they try to escape. The pressure that the molecules of liquid create as they push out into the vapor space is the liquid vapor pressure. More volatile liquids such as LPG have a higher vapor pressure than less volatile diesel oil. **2**. The pressure exerted by a volatile liquid as determined by ASTM D-323, Standard Method of Test for Vapor Pressure of Petroleum Products (Reid Method). **3**. Is a measure of the surface pressure necessary to keep a liquid from vaporizing. The vaporizing tendency of gasoline is measured in terms of its vapor pressure. It is related to vapor lock and engine starting. Vapor lock arises due to the vaporization of the fuel in fuel lines, fuel pump, carburetor, etc., making bubbles of vapor, which prevent the normal flow of fuel. This occurs if the gasoline contains a too-high percentage of low-boiling components as observed by a very high vapor pressure. Alternatively, if the gasoline contains only too few low-boiling components as indicated by a low vapor pressure, then the fuel will not vaporize readily, making it difficult in starting.

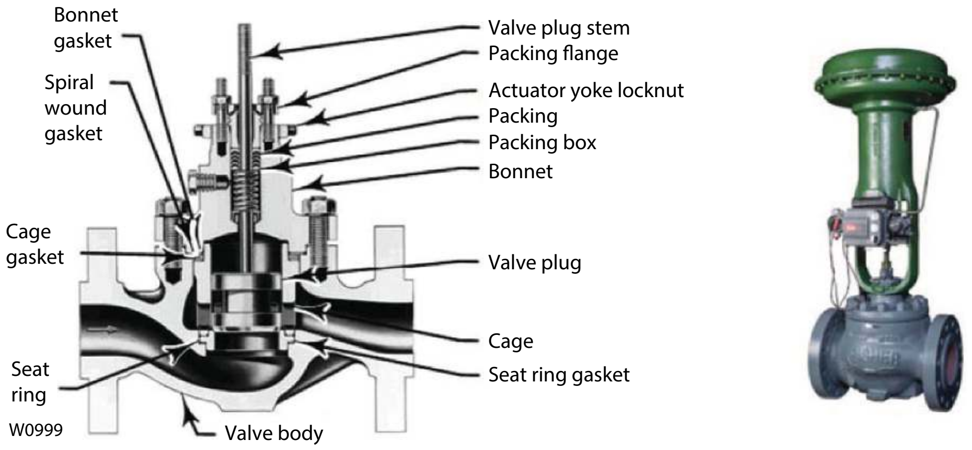


Figure 54 A control valve.

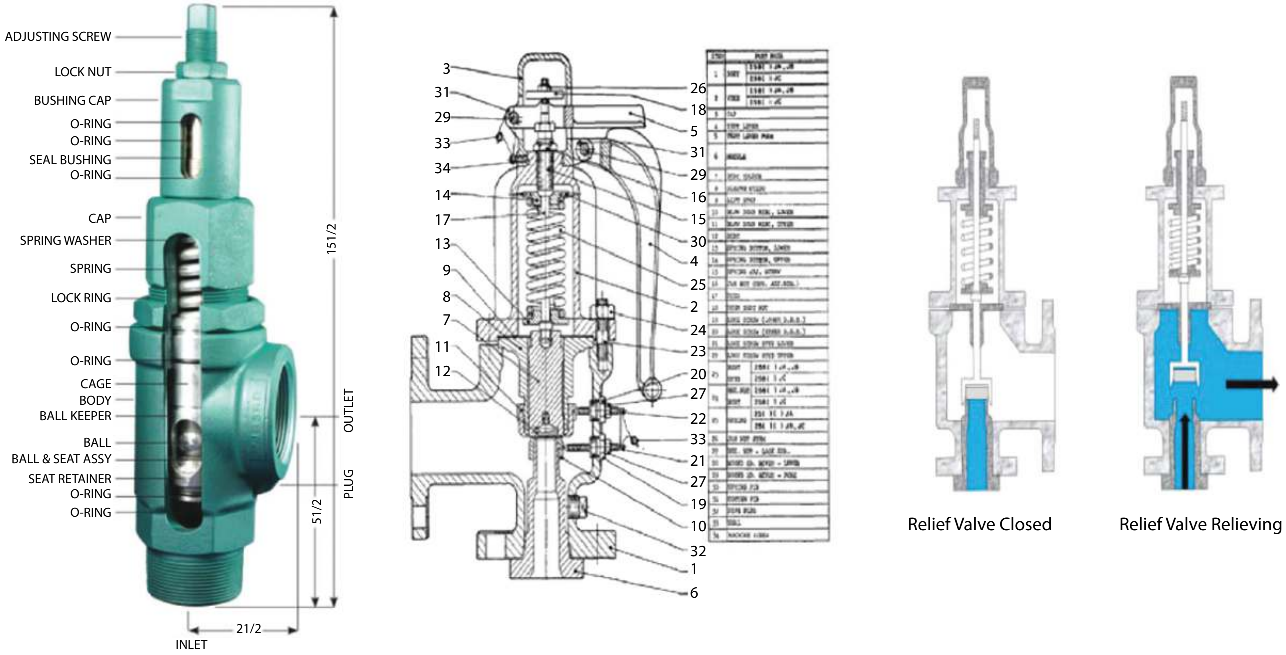


Figure 55 Relief valves.

Vapor Lock: Is the phenomenon of insufficient gasoline flow from a fuel pump due to its inability to pump the mixture that results from low pressure or high temperature, which has high volatility.

Vapor Lock Index: A measure of the tendency of a gasoline to generate excessive vapors in the fuel line, thus causing the displacement of a liquid fuel and a subsequent interruption of normal engine operation. The vapor-lock index ia generally related to RVP and percentage distilled at 158°F (70°C).

Virgin material, gas oil, etc.: Virgin material is material distilled from crude oil but not subjected to processes that chemically alter its composition.

Virgin Stocks: Petroleum oils that have not been cracked or otherwise subjected to any treatment that would produce appreciable chemical change in their components.

Visbreaking: **1**. A thermal cracking process in which heavy atmospheric or vacuum still bottoms are cracked at moderate temperatures to increase the production of distillate products and reduce the viscosity of the distillation residues. **2**. A process in which heavy oil is thermally cracked just enough to lower or break the viscosity. A small quantity of gas oil and lighter products are formed in the process.

Viscosity ASTM D445: The internal resistance to the flow of liquids is expressed as viscosity. The property of liquids under flow conditions that causes them to resist instantaneous change of shape or instantaneous rearrangement of

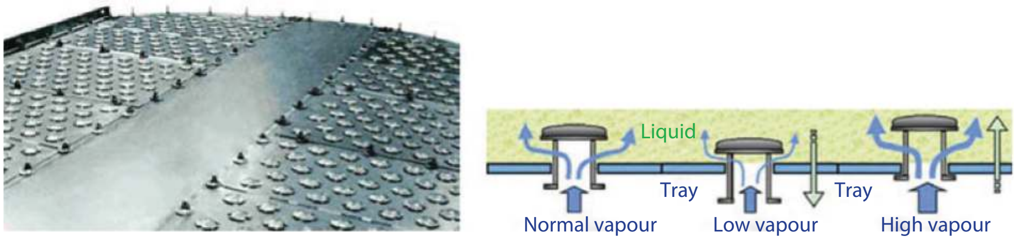


Figure 56 A valve tray.

their parts due to internal friction. Viscosity is generally measured in seconds, at a definite temperature, required for a standard quantity of oil to flow through a standard apparatus. Common viscosity scales in use are Saybolt Universal, Saybolt Furol, poises, kinematic [stokes, or centistokes (cSt)]. Usually, the viscosity measurements are carried out at 100°F (38°C) and 210°F (99°C).

Viscosity is a very important property for the heavy products obtained from the crude oil. The viscosity acts as an important characterization property in the blending units associated to heavy products such as bunker fuel. Typically, the viscosity of these products is specified to be within a specified range, and this is achieved by adjusting the viscosities of the streams entering the blending unit.

Viscosity Index (VI): This index is a series of numbers ranging from 0 to 100 that indicate the rate of change of viscosity with temperature. A Viscosity Index of 100 indicates an oil that does not tend to become viscous at low temperatures or become thin at elevated temperatures.

Typically, paraffin-based lubricating oils exhibit a Viscosity Index of nearly 100, whereas naphthene-based oils on the market show about 40 Viscosity Index, and some naphthenic oils have a Viscosity Index of zero or lower. Paraffin wax has a VI of about 200, and hence its removal reduces the VI of raw lube stocks. By solvent extraction processes, lubricating oils of Viscosity Index higher than 100 can be produced.

Volatile: A hydrocarbon is volatile if it has a sufficient amount of butanes and higher material to noticeably give off vapors at atmospheric conditions.

Volatile Organic Compounds (VOCs): Organic chemicals that have a high vapor pressure at ordinary room temperature. Their high vapor pressure results from a low boiling point, which causes large numbers of molecules to evaporate or sublimate from the liquid or solid form of the compound and enter the surrounding air. For example, formaldehyde (HCHO), which evaporates from paint, has a boiling point of only -19°C (-2°F).

One VOC that is a known human carcinogen is benzene, which is a chemical found in environmental tobacco smoke, stored fuels, and exhaust from cars. Benzene also has natural sources such as volcanoes and forest fires. It is frequently used to make other chemicals in the production of plastics, resins, and synthetic fibers. Benzene evaporates into the air quickly, and the vapor of benzene is heavier than air, allowing the compound to sink into low-lying areas. Benzene has also been known to contaminate food and water and, if digested, can lead to vomiting, dizziness, sleepiness, and rapid heartbeat, and at high levels, even death may occur.

VOCs are many and varied, are dangerous to human health, or cause harm to the environment. Harmful VOCs typically are not acutely toxic but have compounding long-term health effects. Because the concentrations are usually low and the symptoms are slow to develop, the research into VOCs and their effects is difficult.

Volatility: As measured by the distillation characteristics, helps to determine the relative proportion of the various hydrocarbons throughout the boiling range of a gasoline. It is the distillation characteristics along with vapor pressure and vapor/liquid ratio that help to control the performance of the fuel with respect to starting, warm-up, acceleration, vapor-lock, evaporation losses, crankcase dilution, fuel economy, and carburetor icing.

Volatility Factor: An empirical quantity that indicates good gasoline performance with respect to volatility. It involves actual automobile operating conditions and climatic factors. The volatility factor is generally defined as a function of RVP (Reid vapor pressure), percentage distilled at 158°F (70°C), and percentage distilled at 212°F (100°C). This factor is an attempt to predict the vapor-lock tendency of a gasoline.

vppm: Parts per million by volume.

VRC: Vacuum-reduced crude; vacuum tower bottoms.

WABP: Weight average boiling point:

$$\mathrm{WABP} = \sum_{i=1}^{n} X_{wi}\, T_{bi}$$

where

X_{wi} = weight fraction of component i.

T_{bi} = average boiling point of component i.

Wash Zone: A section in a column where the column vapor is washed of entrained heavy materials by contact with a cooler injected liquid. A section of packed material is often used to promote good mixing of the liquid and vapor in the wash zone. All vacuum distillation columns have wash zones to remove heavy residual material that is carried up the column from the flash zone. If washing is inadequate, the heavy residual material forms petroleum coke and plugs the column above the flash zone.

Water Hammer: A violent and potentially damaging shock wave in a pipeline caused by the sudden change in flow rate, such as by the rapid closure of a valve. The effect is avoided by controlling the speed of valve closure, lowering the pressure of the fluid, or lowering the fluid flow rate.

Water vapor: The gaseous state of water dispersed with air at a temperature below the boiling point of the water. The amount present in air is designated by the humidity. The "relative humidity" is the amount of water vapor in a mixture of dry air. A relative humidity of 100% corresponds to the partial pressure of water vapor being equal to the equilibrium vapor pressure and depends on the temperature and pressure.

Watson Characterization Factor (K_w): See Characterization Factor.

Wax: A solid or semi-solid material consisting of a mixture of hydrocarbons obtained or derived from petroleum fractions, or through a Fischer–Tropsch type of process in which the straight- chained paraffins series predominates. This includes all marketable wax, whether crude or refined, with a congealing point (ASTM D 938) between 100 and 200°F (37.8–93°C) and a maximum oil content (ASTM D 3235) of 50% weight.

Weeping: A phenomenon that occurs in a distillation column in which the liquid on a sieve plate passes down through the perforations intended for the vapor to pass up. Weeping occurs when the velocity of the upward vapor is too low. This may be caused by insufficient boil-up.

Weir: A vertical obstruction across a channel carrying a liquid over which the liquid discharges. In a distillation column, a weir is used to retain an amount of liquid on a sieve tray or plate. While the vapor enriched with the more volatile component rises up through the perforations on the sieve tray or plate, the liquid cascades over the weir into the downcomer to the tray below. The weir crest is the top of the weir over which the liquid flows (See Figure 57).

Weighted Average Inlet Temperature (WAIT): [Weight of catalyst in reactor 1 x inlet temperature in reactor 1 + weight of catalyst in reactor 2 x inlet temperature in reactor 2 + weight of catalyst in reactor 3 x inlet temperature in reactors 3]/total weight of catalyst, i.e., [WCR_1 x R_{1IT} + WCR_2 x R_{2IT} + WCR_3 x R_{3IT}]/(WCR_1 + WCR_2 + WCR_3).

where WCR_1, WCR_2, and WCR_3 are the weights of catalysts in reactors 1, 2, and 3, and R_{1IT}, R_{2IT}, and R_{3IT} are the inlet temperatures for reactors 1, 2, and 3, respectively.

Weighted Average Bed Temperature (WABT): [WCR_1 (R_{1IT} + R_{1OT})/2 + WCR_2 (R_{2IT} + R_{2OT})/2 + WCR_3 (R_{3IT} + R_{3OT})/2]/ total weight of catalyst, where WCR_1, WCR_2, and WCR_3 are the weights of the catalyst in reactors 1, 2, and 3;

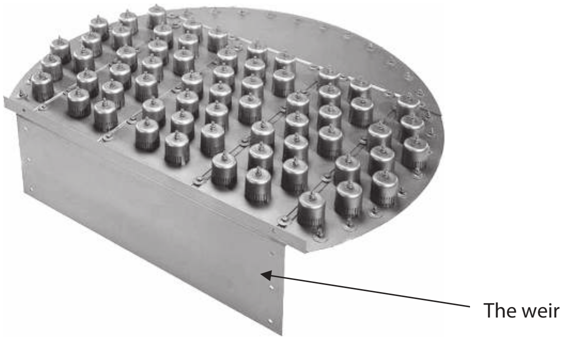


Figure 57 Bubble cap tray with the weir.

R_{1IT}, R_{2IT}, and R_{3IT} are the inlet temperatures for reactors 1, 2, and 3; and R_{1OT}, R_{2OT}, and R_{3OT} are the outlet temperatures for reactors 1, 2, and 3, respectively.

Well: **1**. A natural oil or gas reservoir that exists below a layer of sedimentary rock. **2**. A hole bored or drilled into the earth for the purpose of obtaining water, oil, gas, or other natural resources.

West Texas Intermediate (WTI): A type of crude oil commonly used as a price benchmark.

Wet Gas: **1**. Natural gas that has not had the butane, C_4, and natural gasoline removed. Also the equivalent refinery gas stream. **2**. A term used to describe light hydrocarbon gas dissolved in heavier hydrocarbons. Wet gas is an important source of LPG. **3.** Water that is present in natural gas in offshore platforms. It is necessary to remove the water from the gas for export through subsea pipelines. The pipeline is dosed with corrosion inhibitors to prevent hydrate formation.

White Oil: Sometimes kerosene, sometimes treated kerosene used for pharmaceutical purposes and in the food industry.

WHSV: Weight hour space velocity; weight of feed per hour per weight of a catalyst.

Wick Char: A test used as an indication of the burning quality of a kerosene or illuminating oil. It is defined as the weight of deposits remaining on the wick after a specified amount of sample is burned.

What-If Analysis (WIA): A safety review method, by which "What-If" investigative questions (i.e., brainstorming and/or checklist approach) are asked by an experienced and knowledgeable team of the system or component under review where there are concerns about possible undesired events. Recommendations for the mitigation of identified hazards are provided.

Wppm: Parts per million by weight.

Xylene, $C_6H_4(CH_3)_2$: **1**. A colorless liquid of the aromatic group of hydrocarbons made from the catalytic reforming of certain naphthenic petroleum fractions. Used as high-octane motor and aviation gasoline blending agents, solvents, and chemical intermediates. **2**. One of the aromatic compounds. Xylene has a benzene ring and two methyl radicals attached and three isomers, namely, ortho, para, and metaxylene. Used as a gasoline blending compound or chemical feedstock for making phthalic acids and resins.

Yield: Either the percent of a desired product or all the products resulting from a process involving chemical changes to the feed.

Zeolites: **1**. Compounds used extensively as catalysts, made of silica or aluminum as well as sodium or calcium and other compounds. Zeolites come in a variety of forms—porous and sand like or celatinous—and provide the platform for numerous catalysts. The solid zeolites have extensive pores that give very large surface areas. The precise control during the fabrication of the pore sizes enables selected access to different-sized molecules during reactions. **2**. A class of minerals that are hydrated aluminosilicates. An aluminosilicate is where some of the Si atoms in silica (SiO_4) are replaced with aluminum, giving an excess negative charge. Hydrated means that water is strongly associated with these materials by hydrogen bonding. A positively charged counter ion is required to balance the negative charge on the zeolite. Zeolites are extremely porous materials, with a regular internal structure of cavities of defined size and shape.

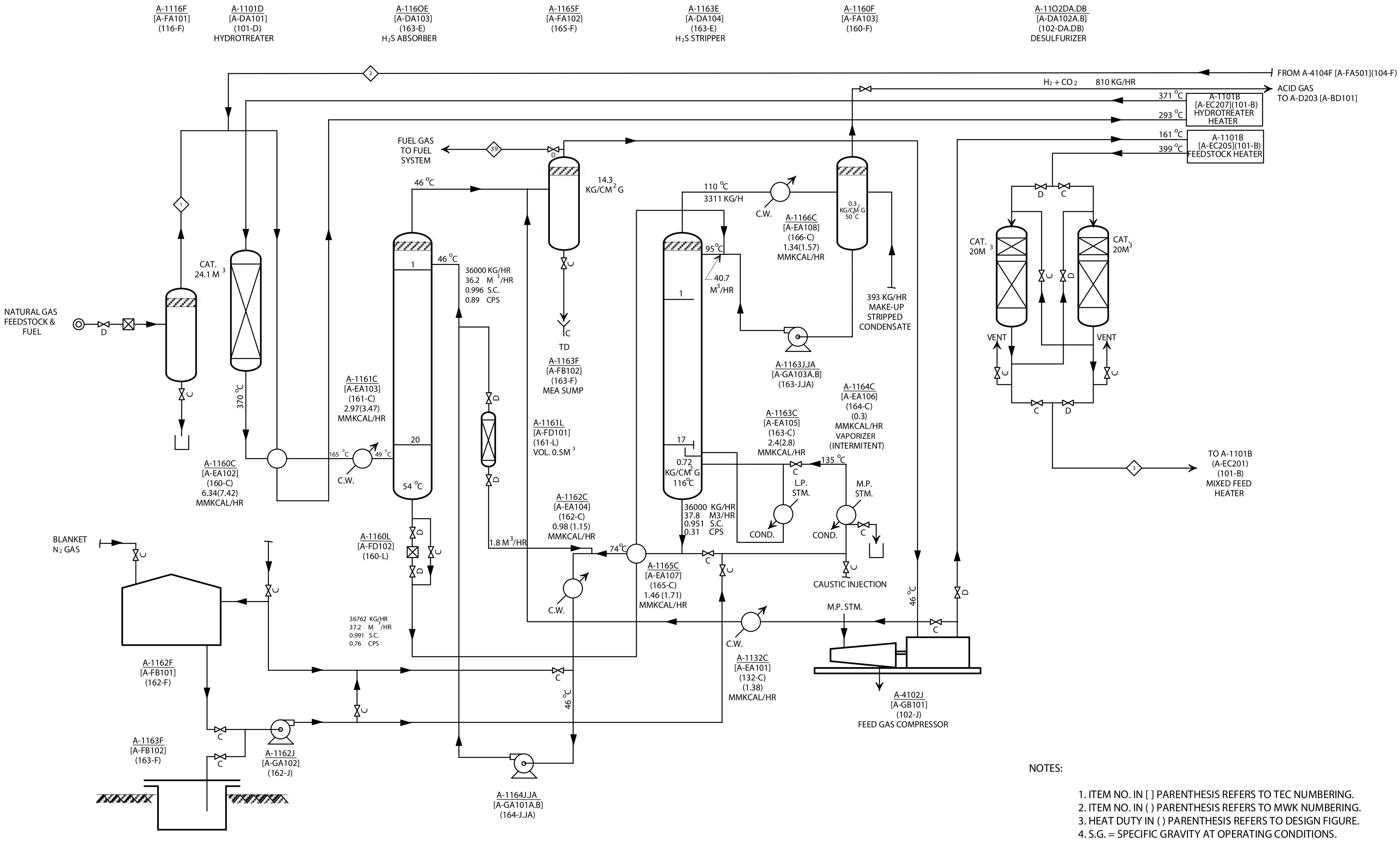
Appendix D

D-1 Process Flow Diagrams Using VISIO 2002 Software

Figure D-1. Process Flow diagram (Feed & Fuel Desulfurization Section).
Figure D-2. Typical process flow diagram for the production of Methyl Tertiary Butyl Ether (MTBE).
Figure D-3. Piping & Instrumentation diagram for Ammonia plant CO_2 removal.
Figure D-4. Piping & Instrumentation diagram (Ammonia synthesis and refrigeration unit).

D-2 Process Data Sheets

1. Air cooled heat exchanger process data sheet
2. Centrifugal pump schedule: driver
3. Centrifugal pump schedule: pump
4. Centrifugal pump summary
5. Column schedule
6. Construction Commissioning Start-up Checklist
7. Deaerator process data sheet: Deaerator water storage tank
8. Deaerator process data sheet: Deaerator head
9. Drum process data sheet
10. Effluent schedule
11. Equilibrium flash calculation
12. Fabricated equipment schedule
13. Fan/Compressor process duty specification
14. Fractionator calculation summary
15. General services and utilities checklist
16. Hazardous chemical and conditions schedule
17. Heat and mass balances
18. Heat exchanger rating sheet
19. Hydrocarbon dew point calculation
20. Line list schedule
21. Line schedule
22. Line schedule sheet
23. Line summary table
24. Mass balance
25. Mechanical equipment schedule
26. Pipe line list
27. Pipe list
28. Piping process conditions summary
29. Plate heat exchanger data sheet
30. Calculation of pressure drop in fixed catalyst beds
31. Process engineering job analysis summary
32. Pump calculation sheet
33. Pump schedule
34. Relief device philosophy sheet
35. Tank and vessel agitator data sheet
36. Tank process data sheet
37. Tank schedule
38. Tie – in – schedule
39. Tower process data sheet
40. Tray loading summary
41. Trip schedule
42. Utility summary sheet
43. Vessel and tank schedule
44. Vessel and tank summary: driver
45. Vessel schedule
46. Water analysis sheet



NOTES:

1. ITEM NO. IN [ ] PARENTHESIS REFERS TO TEC NUMBERING.
2. ITEM NO. IN ( ) PARENTHESIS REFERS TO MWK NUMBERING.
3. HEAT DUTY IN ( ) PARENTHESIS REFERS TO DESIGN FIGURE.
4. S.G. = SPECIFIC GRAVITY AT OPERATING CONDITIONS.
5. CPS = CENTI POISE AT OPERATING CONDITIONS.

Figure D-1. Process Flow diagram (Feed & Fuel Desulfurization Section).

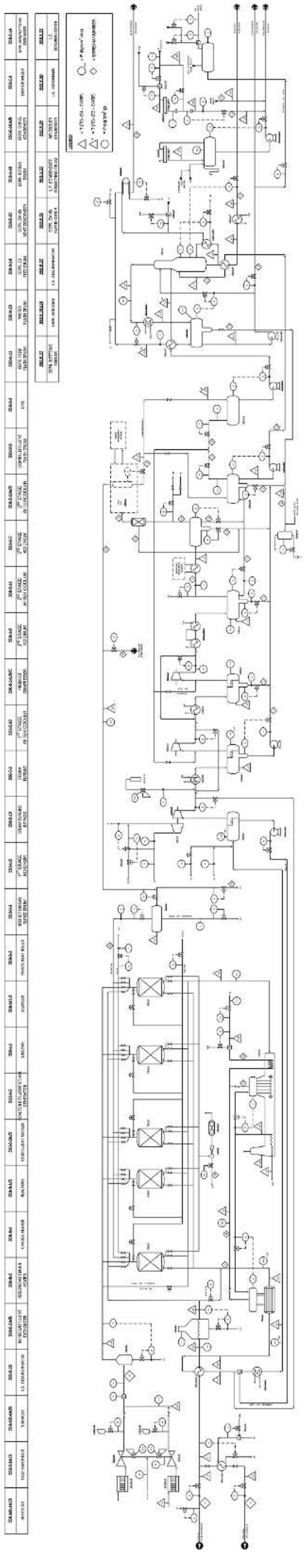


Figure D-2. Typical process flow diagram for the production of Methyl Tertiary Butyl Ether (MTBE).

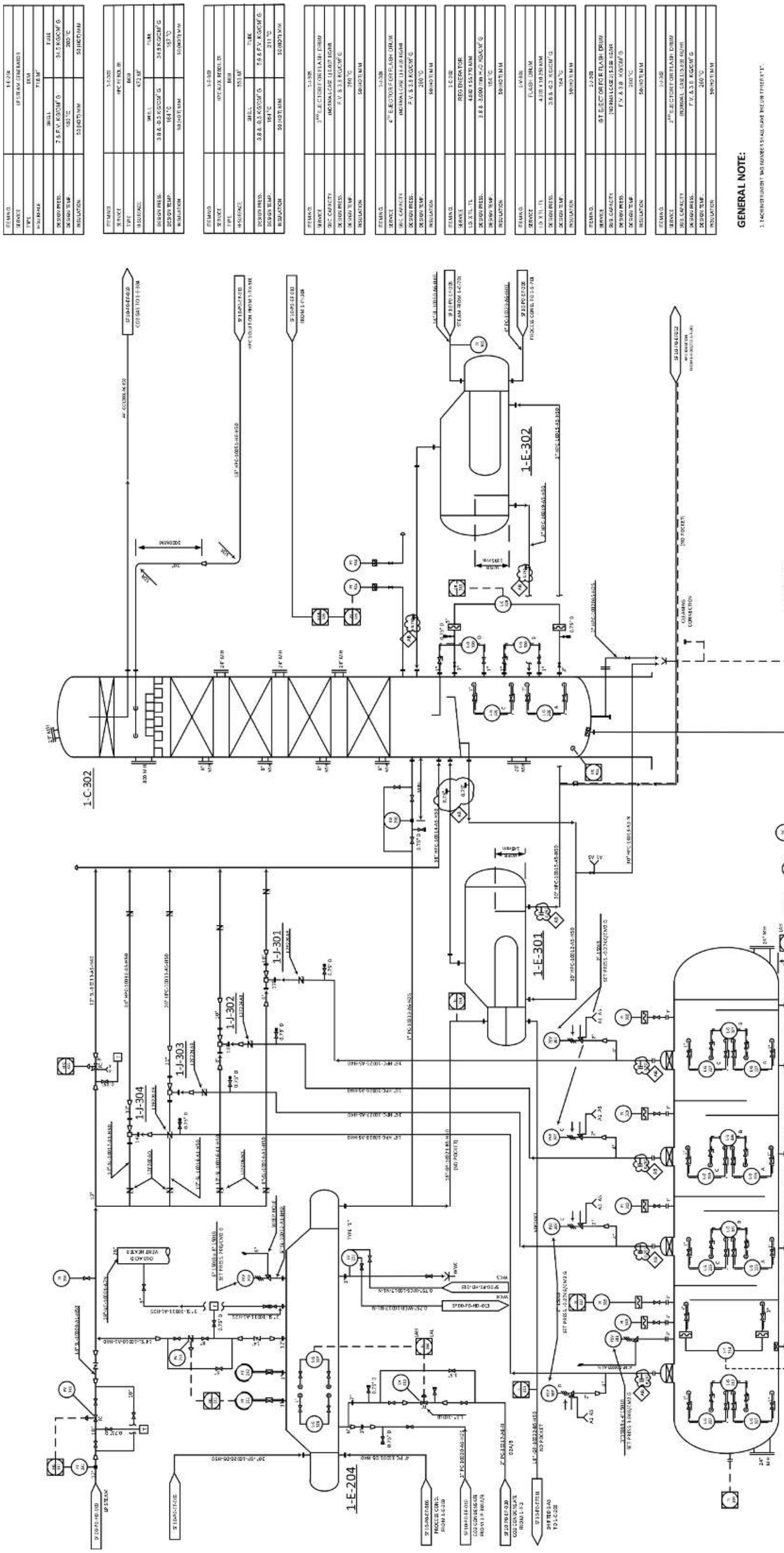
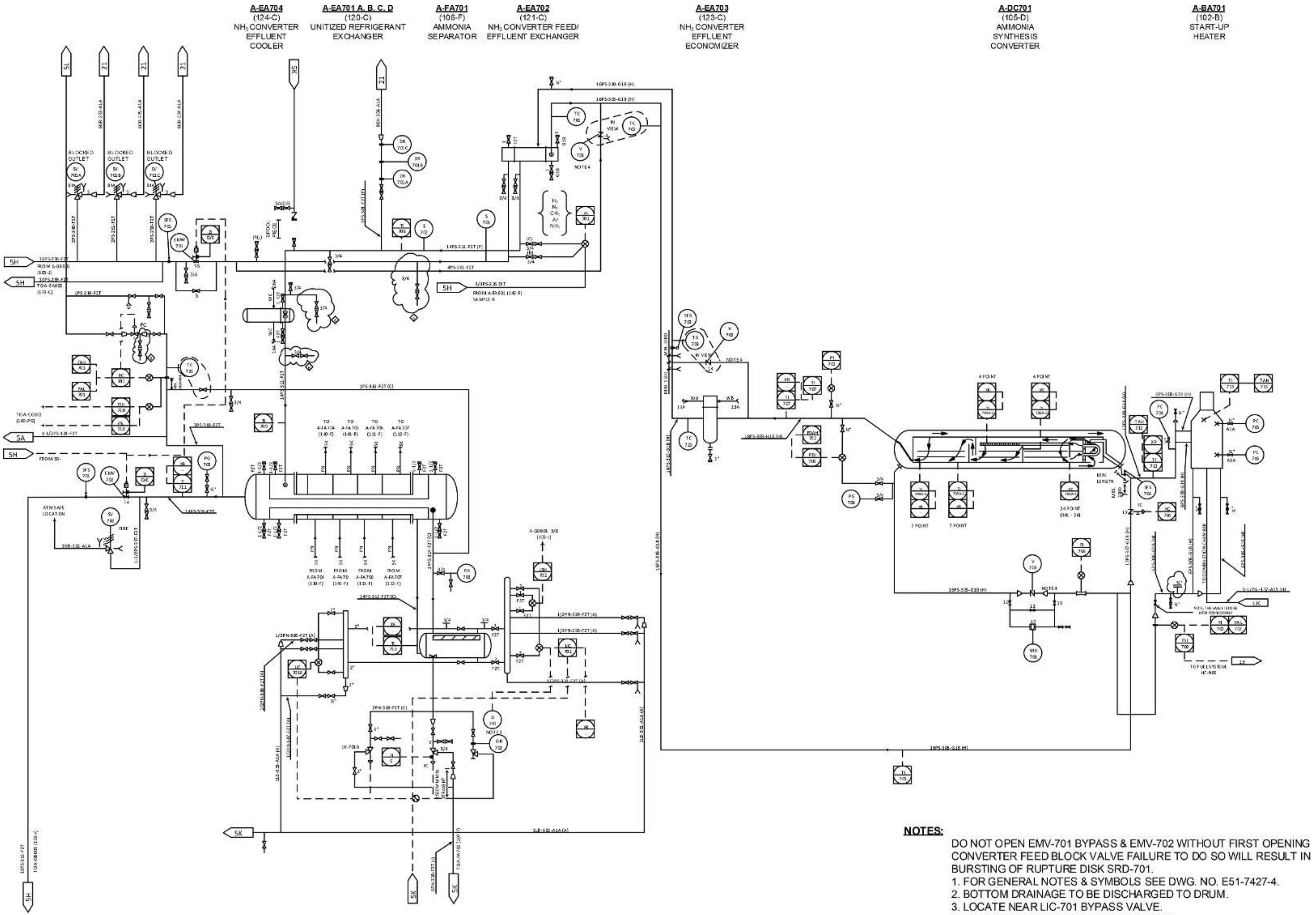


Figure D-3. Piping & Instrumentation diagram for Ammonia plant CO_2 removal.

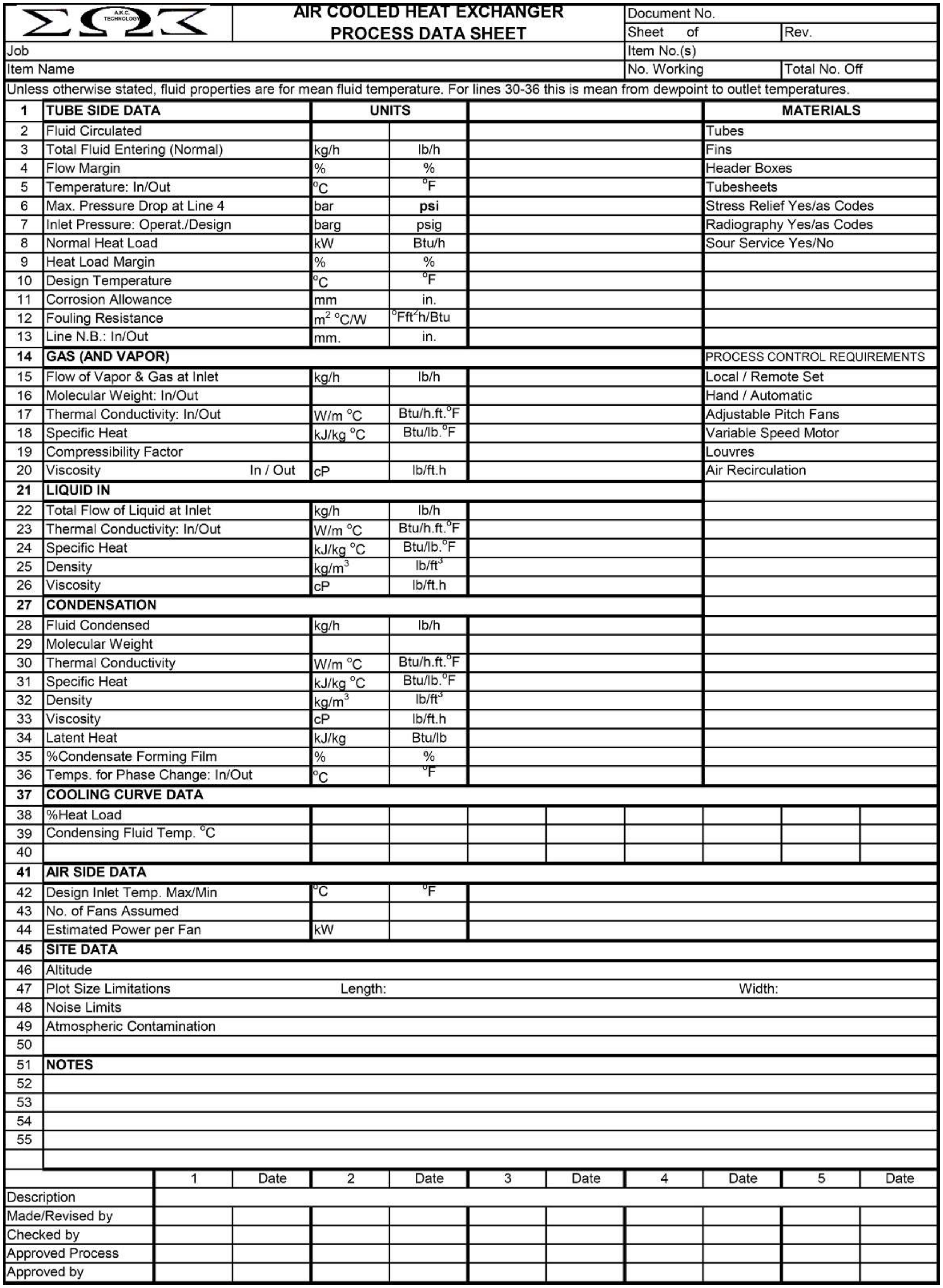


NOTES:

DO NOT OPEN EMV-701 BYPASS & EMV-702 WITHOUT FIRST OPENING CONVERTER FEED BLOCK VALVE FAILURE TO DO SO WILL RESULT IN BURSTING OF RUPTURE DISK SRD-701.

1. FOR GENERAL NOTES & SYMBOLS SEE DWG. NO. E51-7427-4.
2. BOTTOM DRAINAGE TO BE DISCHARGED TO DRUM.
3. LOCATE NEAR LIC-701 BYPASS VALVE.
4. ADJUSTABLE LIMIT STOPPER REQUIRED.

Figure D-4. Piping & Instrumentation flow diagram (Ammonia synthesis and refrigeration unit).



A.K.C. TECHNOLOGY

AIR COOLED HEAT EXCHANGER
PROCESS DATA SHEET

Document No.			
Sheet of		Rev.	
Job		Item No.(s)	
Item Name		No. Working	Total No. Off

Unless otherwise stated, fluid properties are for mean fluid temperature. For lines 30-36 this is mean from dewpoint to outlet temperatures.

1	**TUBE SIDE DATA**	**UNITS**			**MATERIALS**
2	Fluid Circulated				Tubes
3	Total Fluid Entering (Normal)	kg/h	lb/h		Fins
4	Flow Margin	%	%		Header Boxes
5	Temperature: In/Out	°C	°F		Tubesheets
6	Max. Pressure Drop at Line 4	bar	**psi**		Stress Relief Yes/as Codes
7	Inlet Pressure: Operat./Design	barg	psig		Radiography Yes/as Codes
8	Normal Heat Load	kW	Btu/h		Sour Service Yes/No
9	Heat Load Margin	%	%		
10	Design Temperature	°C	°F		
11	Corrosion Allowance	mm	in.		
12	Fouling Resistance	m^2 °C/W	°Fft^{2}h/Btu		
13	Line N.B.: In/Out	mm.	in.		
14	**GAS (AND VAPOR)**				PROCESS CONTROL REQUIREMENTS
15	Flow of Vapor & Gas at Inlet	kg/h	lb/h		Local / Remote Set
16	Molecular Weight: In/Out				Hand / Automatic
17	Thermal Conductivity: In/Out	W/m °C	Btu/h.ft.°F		Adjustable Pitch Fans
18	Specific Heat	kJ/kg °C	Btu/lb.°F		Variable Speed Motor
19	Compressibility Factor				Louvres
20	Viscosity In / Out	cP	lb/ft.h		Air Recirculation
21	**LIQUID IN**				
22	Total Flow of Liquid at Inlet	kg/h	lb/h		
23	Thermal Conductivity: In/Out	W/m °C	Btu/h.ft.°F		
24	Specific Heat	kJ/kg °C	Btu/lb.°F		
25	Density	kg/m^3	lb/ft^3		
26	Viscosity	cP	lb/ft.h		
27	**CONDENSATION**				
28	Fluid Condensed	kg/h	lb/h		
29	Molecular Weight				
30	Thermal Conductivity	W/m °C	Btu/h.ft.°F		
31	Specific Heat	kJ/kg °C	Btu/lb.°F		
32	Density	kg/m^3	lb/ft^3		
33	Viscosity	cP	lb/ft.h		
34	Latent Heat	kJ/kg	Btu/lb		
35	%Condensate Forming Film	%	%		
36	Temps. for Phase Change: In/Out	°C	°F		

37	**COOLING CURVE DATA**								
38	%Heat Load								
39	Condensing Fluid Temp. °C								
40									

41	**AIR SIDE DATA**			
42	Design Inlet Temp. Max/Min	°C	°F	
43	No. of Fans Assumed			
44	Estimated Power per Fan	kW		

45	**SITE DATA**		
46	Altitude		
47	Plot Size Limitations	Length:	Width:
48	Noise Limits		
49	Atmospheric Contamination		
50			
51	**NOTES**		
52			
53			
54			
55			

	1	Date	2	Date	3	Date	4	Date	5	Date
Description										
Made/Revised by										
Checked by										
Approved Process										
Approved by										

CENTRIFUGAL PUMP SCHEDULE
DRIVER

Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
Job Name:			Description						
Document No.			Made/Revised by						
Project No.			Checked by						
Client			Approved- Process						
Location			Approved						
Plant									

ITEM No.	TYPE	H.P.	RPM	ROTATION CW-CCW	ELECTRICAL CHARACTERISTICS	NEMA FRAME	STEAM CONDITIONS		MAKE	MODEL	STEAM RATE	WEIGHT	P.O. No.
							PSIG	TEMP, °F					

Centrifugal pump schedule

CENTRIFUGAL PUMP SCHEDULE

PUMP

Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
Job Name:			Description						
Document No.			Made/Revised by						
Project No.			Checked by						
Client			Approved- Process						
Location			Approved						
Plant									

ITEM No.	No. OF UNITS	SERVICE	MAKE	MODEL	TYPE	SIZE CONNECTION		WEIGHT	REFERENCE DRAWINGS	PURCHASE ORDER No.	SEAL TYPE	TYPE COUPLING	FILE REFERENCE
						SUCTION	DISCHARGE						

Centrifugal pump schedule

CENTRIFUGAL PUMP SUMMARY

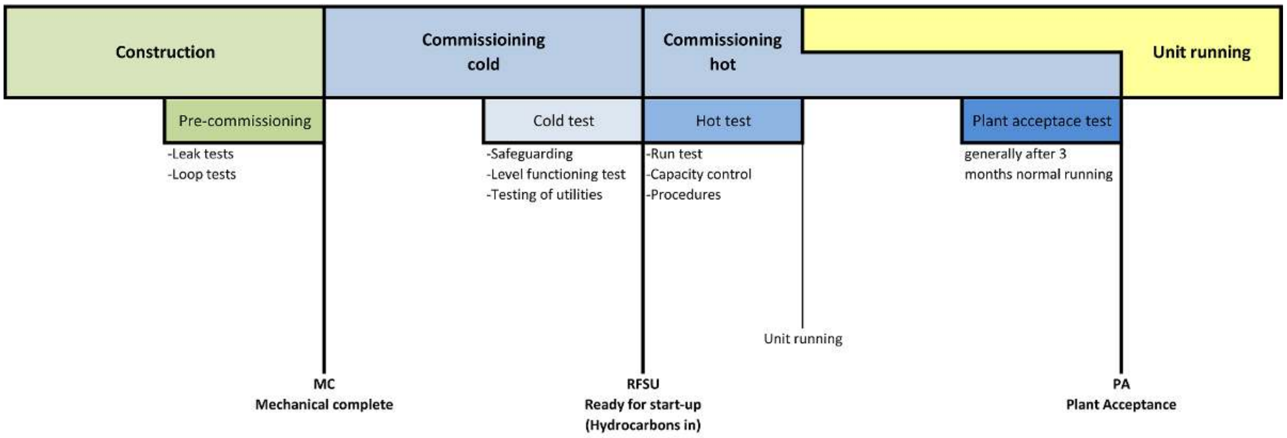
Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
Job Name:			Description						
Document No.			Made/Revised by						
Project No.			Checked by						
Client			Approved- Process						
Location			Approved						
Plant									

ITEM No.	No. OF UNITS	SERVICE	LIQUID	OPERATING TEMPERATURE °F	Sp.Gr OPER. TEMP.	GPM OPER. TEMP.	AVAIL. NPSH ft.	DISCHARGE PRESSURE PSIG	SPEED RPM	BHP PUMP	H.P. DRIVER	DRIVER TYPE	FILE REFERENCE

Centrifugal pump summary

		COLUMN SCHEDULE													
	Job No.:					Sheet of		Issue No.:		1	Date	2	Date	3	Date
	Job Name:							Description							
	Document No.							Made/Revised by							
	Project No.							Checked by							
	Client							Approved- Process							
	Location							Approved							
	Plant														
1	Item Number.														
2	Item Name														
3	Number Required														
4		Units		Top	Bottom	Top	Bottom	Top	Bottom	Top	Bottom	Top	Bottom	Top	Bottom
5	Shell Diameter (I.D. /O.D)	mm													
6	Shell Length (T.L. - T.L.)	mm													
7	Overall Shell Length - (T.L. - T.L.)	mm													
8	Base Elevation	mm													
9	Trays - Type														
10	- Number														
11	- Spacing	mm													
12	- Liquid Flow														
13	Packing - Type														
14	- Number of Beds														
15	- Bed Height / Volume	mm/m^3													
16	- Density	kg/m^3													
17	Other Internals														
18															
19															
20															
21	Operating - Pressure														
22	- Temperature														
23	Design - Pressure	bar g													
24	- Temperature	°C													
25	Vacuum Design														
26	Material: Shell														
27	Liner														
28	Internals														
29															
30	Shell Corrosion Allowance	mm													
31	Sour Service														
32	Stress Relieved for Process Reasons														
33	Insulation														
34															
35															
36	Notes														
37															
38															
39															
40															
41															
42															
43															
44															
45															

Overview Construction - Precommissioining - Commisssioining - Start-up



References

Gen : Installation of Rotating equipment (amendments/supplements to API RP 686)
Reference to installation of rotating equipment
* foundation
* grouting
* flange alignment
* shaft alignment
* rotation check
* oil flushing
* strainers
Reference to commissioning checklist for rotating equipment in general

Gen : Pumps - Type selection and procurement procedure
Reference to selection of pumps including
* data check
* scope vendor
* scope contractor

Gen : Compressors - Selection, testing and installation
Reference to selection of compressors including
* data check
* scope vendor
* scope contractor
* strainers
* air inlet filters

Gen : Centrifugal pumps (amendments/supplements to ISO 13709:2003)
Reference to design of pumps including
* vibration limits
* baseplate design
* lubrication
* nameplate check
* rotation check

Gen : Reciprocating positive displacement pumps and metering pumps (amendments/supplements to API 674 and API 675)
Reference to design of pumps including
* pump dynamics
* baseplate design
* lubrication
* nameplate check
* piston
* crackcase
* dampener

Gen : Axial, centrifugal, and expander compressors (amendments/supplements to API Std 617)
Reference to design of reciprocating compressors including
* vibration limits
* baseplate design
* lubrication
* nameplate check
* seals
* rotation check

Construction Commissioning Start-up Checklist

Gen : Reciprocating compressors (amendments/supplements to API 618) *Reference to design of reciprocating compressors including* * dynamics * baseplate design * lubrication * nameplate check * valves * piston * crackcase * pulsation dampeners
Gen : Rotary-type positive displacement compressors (amendments/supplements to API 619) *Reference to design of positive displacement compressors including* * dynamics * baseplate design * lubrication * nameplate check
Gen : Centrifugal fans (amendments/supplements to ISO 13705, Annex E) *Reference to design of centrifugal fans* * dynamics * baseplate design * lubrication * nameplate check
#NAME?
Gen : Field inspection prior to commissioning of mechanical equipment *Reference to MC checklist for mechanical equipment, for each piece of equipment specific* * rotating equipment * centrifugal compressors * reciprocating compressors * rotary positive displacements pumps * blowers/fans * centrifugal pumps * reciprocating pumps * rotary positive displacement pumps * mixers
Gen : Preservation of new and old equipment standing idle *Reference to preservation of rotating equipment* * preservation oils * cleaniness

Centrifugal pumps - Construction and Pre-commissioning checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Construction - General checks		Remarks	Signed	Signed	Signed	Date
1	Check foundation	Roughen, oil and dust free				
2	Check equipment assembly on foundation	25-50 mm between concrete and base plate, pre-align				
3	Check leveling of base plate					
4	Check installation of foundation bolts					
5	Check equipment for transport damage					
6	Check unit completeness					
7	Check free movement of rotor assembly	E-motor de-energised				

Construction - Checks after grouting		Remarks	Signed	Signed	Signed	Date
8	Check proper grouting is applied	Grout type, sealed anchor bolts, humidity, temperature				
9	Check whether foundations bolts are tightened					
10	Check correct bolts and gaskets are used					
11	Check correct installation coupling					
12	Check coupling guard and mounting					
13	Check cooling water system					
14	Check lubrication system					
15	Check bearings and bearing house are clean					
16	Check oil rings free movement					
17	Check oiler for location and level					
18	Check packing rings in stuffing box					
19	Check mechanical seal					
20	Check seal locks are removed					
21	Check seal liquid system	see installation notes manufacturer				
22	Check equipment for soft feet					
23	Shaft alignment	only metal shims are allowed				
24	Check piping and support attached to pump	Flange deviation				
25	Vents, drain correctly installed					
26	Check earth bosses					
27	Tracing installed correctly					
28	Insulation installed correctly					
29	Check documentation	shaft alignment, foundation measurements, nameplate				
30	P&ID / PEFS check					

Construction - Presommissioning		Remarks	Signed	Signed	Signed	Date
31	Check direction of rotation	uncoupled or avoid mechanical seal runs dry (OH5)				
32	Check N2 pre-charge					
33	Check barrier oil fill					
34	Check lubricant and lubricant levels					

Preservation		Remarks	Signed	Signed	Signed	Date
35	Check equipment for internal cleanliness					
36	Check proper preservation	nozzles blinded off, isolated during flushing				

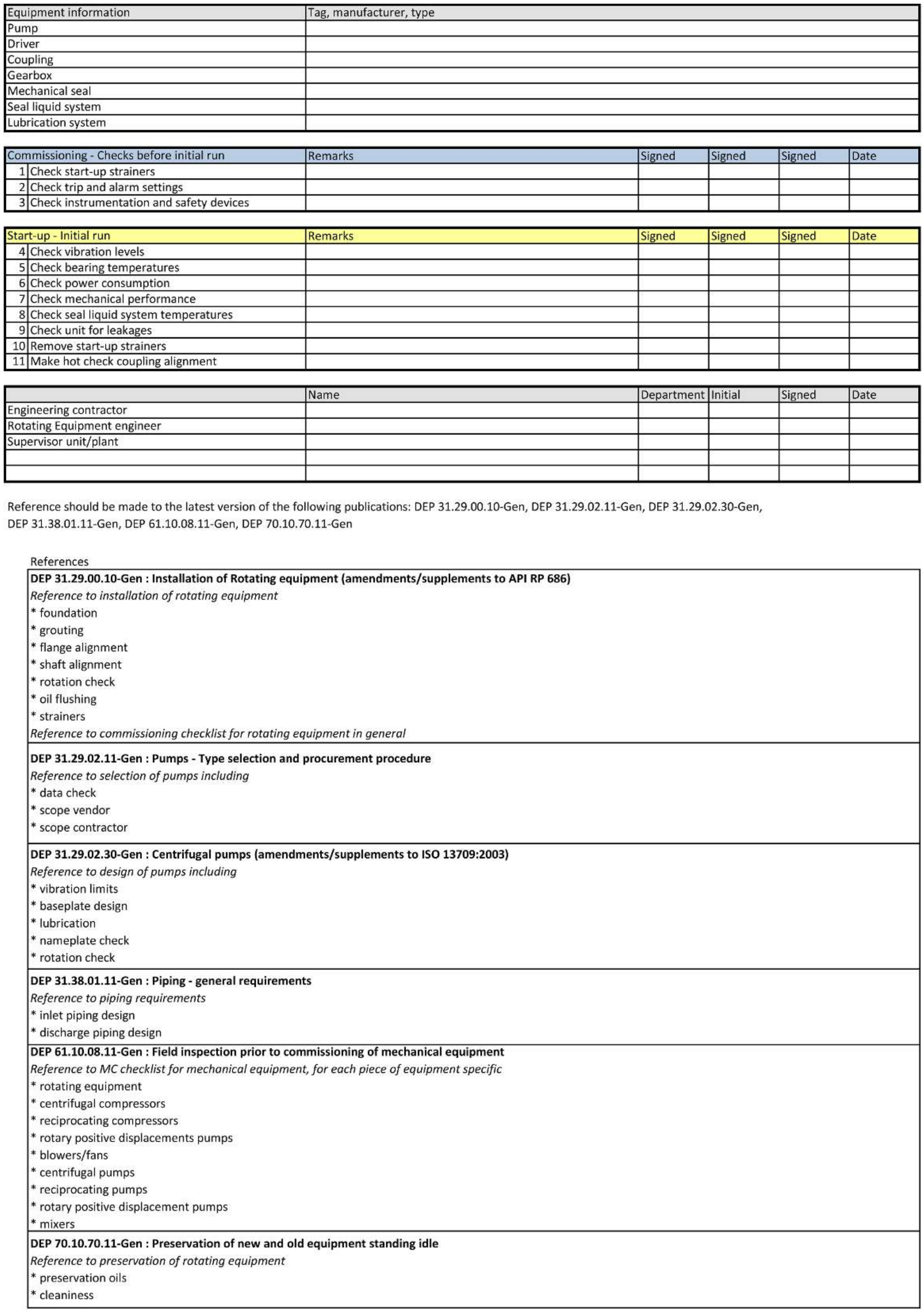
Other items	Remarks
Punch item list attached	

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

Reference should be made to the latest version of the following publications: DEP 31.29.00.10-Gen, DEP 31.29.02.11-Gen, DEP 31.29.02.30-Gen, DEP 31.38.01.11-Gen, DEP 61.10.08.11-Gen, DEP 70.10.70.11-Gen

Centrifugal pumps - Commissioning and Start-up checklist

General information	
Requisition number	
Plant	
Unit	



Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Commissioning - Checks before initial run		Remarks	Signed	Signed	Signed	Date
1	Check start-up strainers					
2	Check trip and alarm settings					
3	Check instrumentation and safety devices					

Start-up - Initial run		Remarks	Signed	Signed	Signed	Date
4	Check vibration levels					
5	Check bearing temperatures					
6	Check power consumption					
7	Check mechanical performance					
8	Check seal liquid system temperatures					
9	Check unit for leakages					
10	Remove start-up strainers					
11	Make hot check coupling alignment					

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

Reference should be made to the latest version of the following publications: DEP 31.29.00.10-Gen, DEP 31.29.02.11-Gen, DEP 31.29.02.30-Gen, DEP 31.38.01.11-Gen, DEP 61.10.08.11-Gen, DEP 70.10.70.11-Gen

References

DEP 31.29.00.10-Gen : Installation of Rotating equipment (amendments/supplements to API RP 686)
Reference to installation of rotating equipment
* foundation
* grouting
* flange alignment
* shaft alignment
* rotation check
* oil flushing
* strainers

Reference to commissioning checklist for rotating equipment in general

DEP 31.29.02.11-Gen : Pumps - Type selection and procurement procedure
Reference to selection of pumps including
* data check
* scope vendor
* scope contractor

DEP 31.29.02.30-Gen : Centrifugal pumps (amendments/supplements to ISO 13709:2003)
Reference to design of pumps including
* vibration limits
* baseplate design
* lubrication
* nameplate check
* rotation check

DEP 31.38.01.11-Gen : Piping - general requirements
Reference to piping requirements
* inlet piping design
* discharge piping design

DEP 61.10.08.11-Gen : Field inspection prior to commissioning of mechanical equipment
Reference to MC checklist for mechanical equipment, for each piece of equipment specific
* rotating equipment
* centrifugal compressors
* reciprocating compressors
* rotary positive displacements pumps
* blowers/fans
* centrifugal pumps
* reciprocating pumps
* rotary positive displacement pumps
* mixers

DEP 70.10.70.11-Gen : Preservation of new and old equipment standing idle
Reference to preservation of rotating equipment
* preservation oils
* cleaniness

Reciprocating pumps - Construction and Pre-commissioning checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Construction - General checks		Remarks	Signed	Signed	Signed	Date
1	Check foundation	Roughen, oil and dust free				
2	Check equipment assembly on foundation	25-50 mm between concrete and base plate, pre-align				
3	Check leveling of base plate					
4	Check installation of foundation bolts					
5	Check equipment for transport damage					
6	Check unit completeness					
7	Check free movement of piston/ motor rotor	E-motor de-energised				

Construction - Checks after grouting		Remarks	Signed	Signed	Signed	Date
8	Check proper grouting is applied	Grout type, sealed anchor bolts, humidity, temperature				
9	Check whether foundations bolts are tightened					
10	Check correct bolts and gaskets are used					
11	Check crack case for cleanliness					
12	Check suction and discharge valve assembly					
13	Check tension of V-belts and pulley alignment					
14	Check valve gear					
15	Check pulsation dampeners					
16	Check correct installation coupling					
17	Check coupling guard and mounting					
18	Check cooling water system					
19	Check lubrication system					
20	Check bearings and bearing house are clean					
21	Check stuffing box materials					
22	Check seal					
23	Check seal liquid system	see installation notes manufacturer				
24	Check equipment for soft feet					
25	Shaft alignment	only metal shims are allowed				
26	Check piping and support attached to pump	Flange deviation				
27	Vents, drain correctly installed					
28	Check earth bosses					
29	Tracing installed correctly					
30	Insulation installed correctly					
31	Check documentation	shaft alignment, foundation measurements, nameplate				
32	P&ID / PEFS check					

Construction - Presommissioning		Remarks	Signed	Signed	Signed	Date
33	Check direction of rotation	uncoupled or avoid seal/bearing runs dry (OH5)				
34	Check N2 pre-charge					
35	Check barrier oil fill					
36	Check lubricant and lubricant levels					

Preservation		Remarks	Signed	Signed	Signed	Date
37	Check equipment for internal cleanliness					
38	Check proper preservation	nozzles blinded off, isolated during flushing				

Other items	Remarks
Punch item list attached	

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

Reference should be made to the latest version of the following publications: DEP 31.29.00.10-Gen, DEP 31.29.02.11-Gen, DEP 31.29.12.30-Gen, DEP 31.38.01.11-Gen, DEP61.10.08.11-Gen, DEP 70.10.70.11-Gen

Reciprocating pumps - Commissioning and Start-up checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Commissioning - Checks before initial run		Remarks	Signed	Signed	Signed	Date
1	Check start-up strainers					
2	Check trip and alarm settings					
3	Check instrumentation and safety devices					

Start-up - Initial run		Remarks	Signed	Signed	Signed	Date
4	Check vibration levels					
5	Check bearing temperatures					
6	Check power consumption					
7	Check mechanical performance					
8	Check proper functioning pulsation dampeners					
9	Check seal or stuffing box temperatures					
10	Check unit for leakages					
11	Remove start-up strainers					
12	Make hot check coupling alignment					

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

Reference should be made to the latest version of the following publications: DEP 31.29.00.10-Gen, DEP 31.29.02.11-Gen, DEP 31.29.12.30-Gen, DEP 31.38.01.11-Gen, DEP61.10.08.11-Gen, DEP 70.10.70.11-Gen

References

DEP 31.29.00.10-Gen : Installation of Rotating equipment (amendments/supplements to API RP 686)

Reference to installation of rotating equipment

* foundation
* grouting
* flange alignment
* shaft alignment
* rotation check
* oil flushing
* strainers

Reference to commissioning checklist for rotating equipment in general

DEP 31.29.02.11-Gen : Pumps - Type selection and procurement procedure

Reference to selection of pumps including

* data check
* scope vendor
* scope contractor

DEP 31.29.12.30-Gen : Reciprocating positive displacement pumps and metering pumps (amendments/supplements to API 674 and API 675)

Reference to design of pumps including

* pump dynamics
* baseplate design
* lubrication
* nameplate check
* piston
* crackcase
* dampener

DEP 31.38.01.11-Gen : Piping - general requirements

Reference to piping requirements

* inlet piping design
* discharge piping design

DEP 61.10.08.11-Gen : Field inspection prior to commissioning of mechanical equipment

Reference to MC checklist for mechanical equipment, for each piece of equipment specific

* rotating equipment
* centrifugal compressors
* reciprocating compressors
* rotary positive displacements pumps
* blowers/fans
* centrifugal pumps
* reciprocating pumps
* rotary positive displacement pumps
* mixers

DEP 70.10.70.11-Gen : Preservation of new and old equipment standing idle

Reference to preservation of rotating equipment

* preservation oils
* cleaniness

Rotary positive displacement pumps - Construction and Pre-commissioning checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Construction - General checks		Remarks	Signed	Signed	Signed	Date
1	Check foundation	Roughen, oil and dust free				
2	Check equipment assembly on foundation	25-50 mm between concrete and base plate, pre-align				
3	Check leveling of base plate					
4	Check installation of foundation bolts					
5	Check equipment for transport damage					
6	Check unit completeness					
7	Check free movement of rotor assembly	E-motor de-energised				

Construction - Checks after grouting		Remarks	Signed	Signed	Signed	Date
8	Check proper grouting is applied	Grout type, sealed anchor bolts, humidity, temperature				
9	Check whether foundations bolts are tightened					
10	Check correct bolts and gaskets are used					
11	Check correct installation coupling					
12	Check coupling guard and mounting					
13	Check cooling water system					
14	Check lubrication system					
15	Check bearings and bearing house are clean					
16	Check oil rings free movement					
17	Check oiler for location and level					
18	Check packing rings in stuffing box					
19	Check mechanical seal					
20	Check cleanliness internals					
21	Check seal liquid system	see installation notes manufacturer				
22	Check equipment for soft feet					
23	Shaft alignment	only metal shims are allowed				
24	Check piping and support attached to pump	Flange deviation				
25	Vents, drain correctly installed					
26	Check earth bosses					
27	Tracing installed correctly					
28	Insulation installed correctly					
29	Check documentation	shaft alignment, foundation measurements, nameplate				
30	P&ID / PEFS check					

Construction - Presommissioning		Remarks	Signed	Signed	Signed	Date
31	Check direction of rotation	uncoupled or avoid dry run				
32	Check N2 pre-charge					
33	Check barrier oil fill					
34	Check lubricant and lubricant levels					

Preservation		Remarks	Signed	Signed	Signed	Date
35	Check equipment for internal cleanliness					
36	Check proper preservation	nozzles blinded off, isolated during flushing				

Other items	Remarks
Punch item list attached	

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

Rotary positive displacement pumps - Commissioning and Start-up checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Commissioning - Checks before initial run		Remarks	Signed	Signed	Signed	Date
1	Check start-up strainers					
2	Check trip and alarm settings					
3	Check instrumentation and safety devices					

Start-up - Initial run		Remarks	Signed	Signed	Signed	Date
4	Check vibration levels					
5	Check bearing temperatures					
6	Check power consumption					
7	Check mechanical performance					
8	Check seal liquid system temperatures					
9	Check liquid has sufficient lubricating properties					
10	Check unit for leakages					
11	Remove start-up strainers					
12	Make hot check coupling alignment					

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

References

Gen : Installation of Rotating equipment (amendments/supplements to API RP 686)

Reference to installation of rotating equipment

* foundation
* grouting
* flange alignment
* shaft alignment
* rotation check
* oil flushing
* strainers

Reference to commissioning checklist for rotating equipment in general

Gen : Pumps - Type selection and procurement procedure

Reference to selection of pumps including

* data check
* scope vendor
* scope contractor

Gen : Piping - general requirements

Reference to piping requirements

* inlet piping design
* discharge piping design

Gen : Field inspection prior to commissioning of mechanical equipment

Reference to MC checklist for mechanical equipment, for each piece of equipment specific

* rotating equipment
* centrifugal compressors
* reciprocating compressors
* rotary positive displacements pumps
* blowers/fans
* centrifugal pumps
* reciprocating pumps
* rotary positive displacement pumps
* mixers

Gen : Preservation of new and old equipment standing idle

Reference to preservation of rotating equipment

* preservation oils
* cleaniness

Centrifugal Compressors - Construction and Pre-commissioning checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Construction - General checks		Remarks	Signed	Signed	Signed	Date
1	Check foundation	Roughen, oil and dust free				
2	Check equipment assembly on foundation					
3	Check leveling of base plate					
4	Check installation of foundation bolts					
5	Check equipment for transport damage					
6	Check unit completeness					
7	Check free movement of rotor assembly	E-motor de-energised				

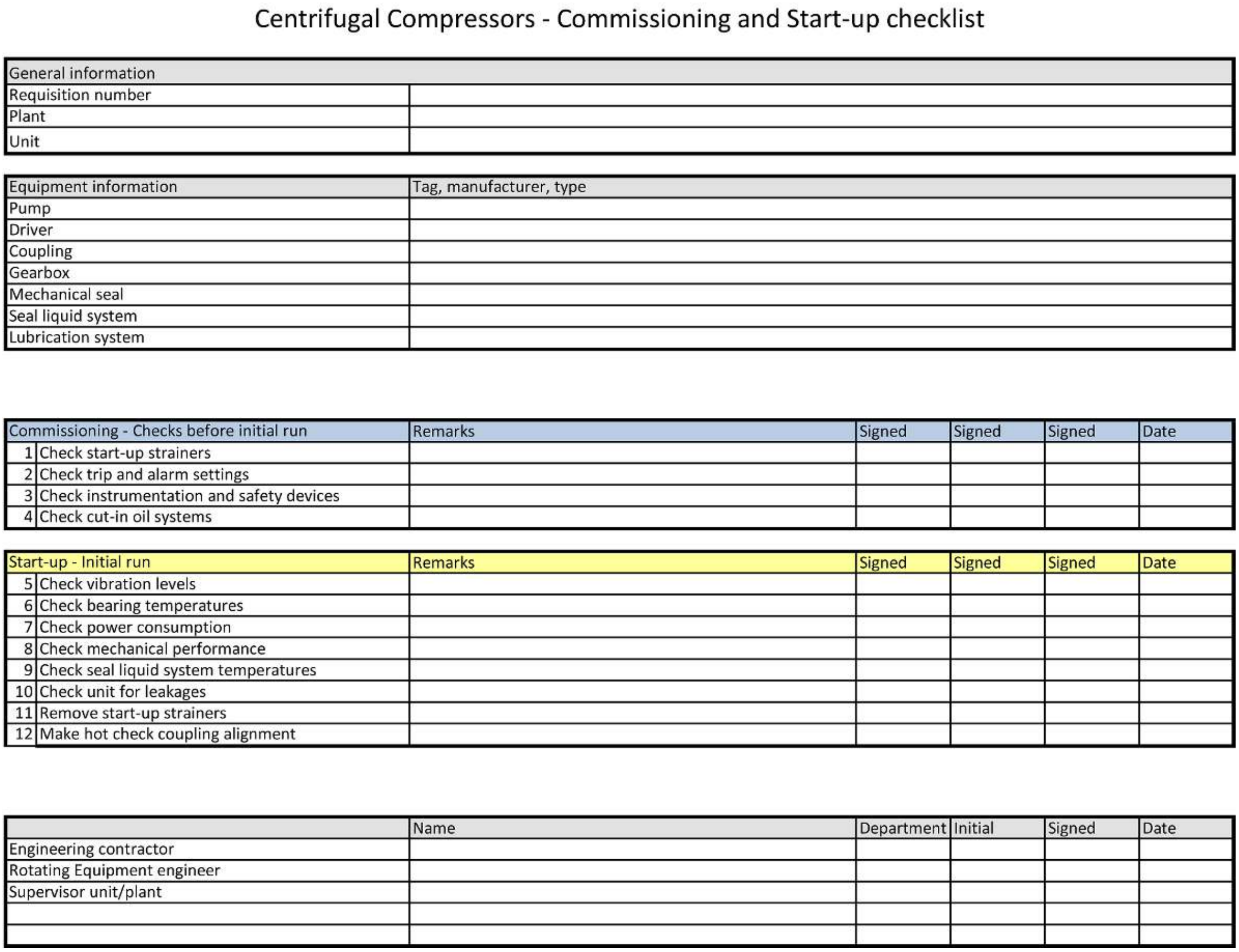
Construction - Checks after grouting		Remarks	Signed	Signed	Signed	Date
8	Check proper grouting is applied	Grout type, sealed anchor bolts, humidity, temperature				
9	Check whether foundations bolts are tightened					
10	Check correct bolts and gaskets are used					
11	Check correct installation coupling					
12	Check coupling guard and mounting					
13	Check cooling water system					
14	Check lubrication system					
15	Check cleaniness lube oil tank					
16	Check bearings and bearing house are clean					
17	Check oil rings free movement					
18	Check functioning sour oil traps					
19	Check casing can expand					
20	Check mechanical seal					
21	Check seal locks are removed					
22	Check seal liquid system	see installation notes manufacturer				
23	Check proper functioning barring gear					
24	Check axial displacement rotor					
25	Check equipment for soft feet					
26	Shaft alignment	only metal shims are allowed				
27	Check piping and support attached to pump	Flange deviation				
28	Vents, drain correctly installed					
29	Check earth bosses					
30	Tracing installed correctly					
31	Insulation installed correctly					
32	Check documentation	shaft alignment, foundation measurements, nameplate				
33	P&ID / PEFS check					

Construction - Presommissioning		Remarks	Signed	Signed	Signed	Date
34	Check direction of rotation					
35	Check N2 pre-charge					
36	Check barrier oil fill					
37	Check lubricant and lubricant levels					

Preservation		Remarks	Signed	Signed	Signed	Date
38	Check equipment for internal cleanliness					
39	Check proper preservation	nozzles blinded off, isolated during flushing				

Other items	Remarks
Punch item list attached	

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					



Centrifugal Compressors - Commissioning and Start-up checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Commissioning - Checks before initial run		Remarks	Signed	Signed	Signed	Date
1	Check start-up strainers					
2	Check trip and alarm settings					
3	Check instrumentation and safety devices					
4	Check cut-in oil systems					

Start-up - Initial run		Remarks	Signed	Signed	Signed	Date
5	Check vibration levels					
6	Check bearing temperatures					
7	Check power consumption					
8	Check mechanical performance					
9	Check seal liquid system temperatures					
10	Check unit for leakages					
11	Remove start-up strainers					
12	Make hot check coupling alignment					

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

Gen : Installation of Rotating equipment (amendments/supplements to API RP 686)
Reference to installation of rotating equipment
* foundation
* grouting
* flange alignment
* shaft alignment
* rotation check
* oil flushing
* strainers
Reference to commissioning checklist for rotating equipment in general

Gen : Compressors - Selection, testing and installation
Reference to selection of compressors including
* data check
* scope vendor
* scope contractor
* strainers
* air inlet filters

Gen : Axial, centrifugal, and expander compressors (amendments/supplements to API Std 617)
Reference to design of reciprocating compressors including
* vibration limits
* baseplate design
* lubrication
* nameplate check
* seals
* rotation check

Gen : Piping - general requirements
Reference to piping requirements
* inlet piping design
* discharge piping design

Gen : Field inspection prior to commissioning of mechanical equipment
Reference to MC checklist for mechanical equipment, for each piece of equipment specific
* rotating equipment
* centrifugal compressors
* reciprocating compressors
* rotary positive displacements pumps
* blowers/fans
* centrifugal pumps
* reciprocating pumps
* rotary positive displacement pumps
* mixers

Gen : Preservation of new and old equipment standing idle
Reference to preservation of rotating equipment
* preservation oils
* cleaniness

Gen : Preservation of new and old equipment standing idle
Reference to preservation of rotating equipment
* preservation oils
* cleaniness

Reciprocating Compressors - Construction and Pre-commissioning checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Construction - General checks		Remarks	Signed	Signed	Signed	Date
1	Check foundation	Roughen, oil and dust free				
2	Check equipment assembly on foundation	25-50 mm between concrete and base plate, pre-align				
3	Check leveling of base plate					
4	Check installation of foundation bolts					
5	Check equipment for transport damage					
6	Check unit completeness					
7	Check alignment crosshead guides					

Construction - Checks after grouting		Remarks	Signed	Signed	Signed	Date
8	Check proper grouting is applied	Grout type, sealed anchor bolts, humidity, temperature				
9	Check whether foundations bolts are tightened					
10	Check correct bolts and gaskets are used					
11	Check crackshaft deflection					
12	Check suction and discharge valve assembly					
13	Check piston travel clearance					
14	Check pulsation dampeners					
15	Check funtioning clearance pockets					
16	Check tension and alignment of V-belt					
17	Check support flywheel					
18	Check cilinder mounting and alignment					
19	Check correct installation coupling					
20	Check coupling guard and mounting					
21	Check cooling water system					
22	Check lubrication system					
23	Inspect lube oil tank					
24	Inspect lubricator					
25	Check seal					
26	Check seal purch					
27	Check crackcase and crosshead cleanliness					
28	Check proper functioning barring gear					
29	Check equipment for soft feet					
30	Shaft alignment	only metal shims are allowed				
31	Check piping and support attached to pump	Flange deviation				
32	Vents, drain correctly installed					
33	Check earth bosses					
34	Tracing installed correctly					
35	Insulation installed correctly					
36	Check documentation	shaft alignment, foundation measurements, nameplate				
37	P&ID / PEFS check					

Construction - Presommissioning		Remarks	Signed	Signed	Signed	Date
38	Check direction of rotation	(jacking oil pump)				
39	Check N2 pre-charge					
40	Check barrier oil fill					
41	Check lubricant and lubricant levels					

Preservation		Remarks	Signed	Signed	Signed	Date
42	Check equipment for internal cleanliness					
43	Check proper preservation	nozzles blinded off, isolated during flushing				

Other items	Remarks
Punch item list attached	

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

Reciprocating Compressors - Commissioning and Start-up checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Commissioning - Checks before initial run		Remarks	Signed	Signed	Signed	Date
1	Check start-up strainers					
2	Check trip and alarm settings					
3	Check instrumentation and safety devices					

Start-up - Initial run		Remarks	Signed	Signed	Signed	Date
4	Check vibration levels					
5	Check bearing temperatures					
6	Check power consumption					
7	Check mechanical performance					
8	Check seal or stuffing box temperatures					
9	Check loading/unloading devices					
10	Check unit for leakages					
11	Remove start-up strainers					
12	Make hot check coupling alignment					

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

References

Gen : Installation of Rotating equipment (amendments/supplements to API RP 686) *Reference to installation of rotating equipment* * foundation * grouting * flange alignment * shaft alignment * rotation check * oil flushing * strainers *Reference to commissioning checklist for rotating equipment in general*
Gen : Compressors - Selection, testing and installation *Reference to selection of compressors including* * data check * scope vendor * scope contractor * strainers * air inlet filters
Gen : Reciprocating compressors (amendments/supplements to API 618) *Reference to design of reciprocating compressors including* * dynamics * baseplate design * lubrication * nameplate check * valves * piston * crackcase * pulsation dampeners
Gen : Piping - general requirements *Reference to piping requirements* * inlet piping design * discharge piping design
Gen : Field inspection prior to commissioning of mechanical equipment *Reference to MC checklist for mechanical equipment, for each piece of equipment specific* * rotating equipment * centrifugal compressors * reciprocating compressors * rotary positive displacements pumps * blowers/fans * centrifugal pumps * reciprocating pumps * rotary positive displacement pumps * mixers
Gen : Preservation of new and old equipment standing idle *Reference to preservation of rotating equipment* * preservation oils * cleaniness

Screw Compressors - Construction and Pre-commissioning checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Construction - General checks		Remarks	Signed	Signed	Signed	Date
1	Check foundation	Roughen, oil and dust free				
2	Check equipment assembly on foundation	25-50 mm between concrete and base plate, pre-align				
3	Check leveling of base plate					
4	Check installation of foundation bolts					
5	Check equipment for transport damage					
6	Check unit completeness					

Construction - Checks after grouting		Remarks	Signed	Signed	Signed	Date
7	Check proper grouting is applied	Grout type, sealed anchor bolts, humidity, temperature				
8	Check whether foundations bolts are tightened					
9	Check correct bolts and gaskets are used					
10	Check correct installation coupling					
11	Check coupling guard and mounting					
12	Check internals for cleanliness					
13	Check suction line for cleanliness					
14	Check tention and alignment V-belt					
15	Check cooling water system					
16	Check lubrication system					
17	Check bearings and bearing house are clean					
18	Check seal					
19	Check seal liquid system	see installation notes manufacturer				
20	Check equipment for soft feet					
21	Shaft alignment	only metal shims are allowed				
22	Check piping and support attached to pump	Flange deviation				
23	Vents, drain correctly installed					
24	Check earth bosses					
25	Tracing installed correctly					
26	Insulation installed correctly					
27	Check documentation	shaft alignment, foundation measurements, nameplate				
28	P&ID / PEFS check					

Construction - Presommissioning		Remarks	Signed	Signed	Signed	Date
29	Check direction of rotation	uncoupled or avoid mechanical seal runs dry				
30	Check N2 pre-charge					
31	Check barrier oil fill					
32	Check lubricant and lubricant levels					

Preservation		Remarks	Signed	Signed	Signed	Date
33	Check equipment for internal cleanliness					
34	Check proper preservation	nozzles blinded off, isolated during flushing				

Other items	Remarks
Punch item list attached	

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

Screw Compressors - Commissioning and Start-up checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Commissioning - Checks before initial run		Remarks	Signed	Signed	Signed	Date
1	Check start-up strainers					
2	Check trip and alarm settings					
3	Check instrumentation and safety devices					

Start-up - Initial run		Remarks	Signed	Signed	Signed	Date
4	Check vibration levels					
5	Check bearing temperatures					
6	Check power consumption					
7	Check mechanical performance					
8	Check seal liquid system temperatures					
9	Check unit for leakages					
10	Remove start-up strainers					
11	Make hot check coupling alignment					

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

References
Gen : Installation of Rotating equipment (amendments/supplements to API RP 686) *Reference to installation of rotating equipment* * foundation * grouting * flange alignment * shaft alignment * rotation check * oil flushing * strainers *Reference to commissioning checklist for rotating equipment in general*
Gen : Compressors - Selection, testing and installation *Reference to selection of compressors including* * data check * scope vendor * scope contractor * strainers * air inlet filters
Gen : Rotary-type positive displacement compressors (amendments/supplements to API 619) *Reference to design of positive displacement compressors including* * dynamics * baseplate design * lubrication * nameplate check
Gen : Piping - general requirements *Reference to piping requirements* * inlet piping design * discharge piping design
#NAME?
Gen : Preservation of new and old equipment standing idle *Reference to preservation of rotating equipment* * preservation oils * cleaniness

Blowers-Fans - Construction and Pre-commissioning checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Construction - General checks		Remarks	Signed	Signed	Signed	Date
1	Check foundation	Roughen, oil and dust free				
2	Check equipment assembly on foundation	25-50 mm between concrete and base plate, pre-align				
3	Check leveling of base plate					
4	Check installation of foundation bolts					
5	Check equipment for transport damage					
6	Check unit completeness					
7	Check free movement of rotor assembly	E-motor de-energised				

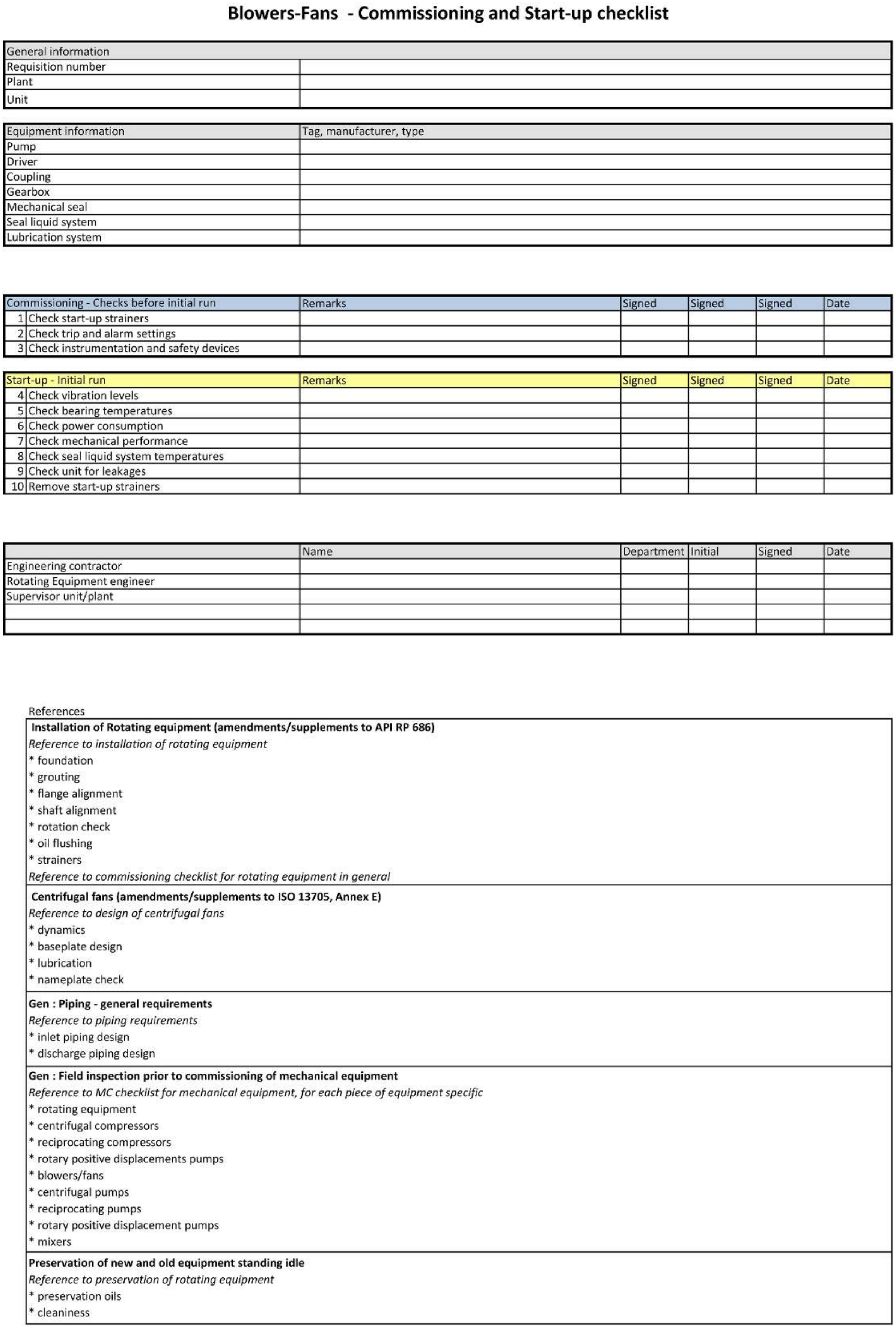
Construction - Checks after grouting		Remarks	Signed	Signed	Signed	Date
8	Check proper grouting is applied	Grout type, sealed anchor bolts, humidity, temperature				
9	Check whether foundations bolts are tightened					
10	Check correct bolts and gaskets are used					
11	Check radial and axial clearance impeller-volute					
12	Check correct installation coupling					
13	Check coupling guard and mounting					
14	Check cooling water system					
15	Check lubrication system					
16	Check bearings and bearing house are clean					
17	Check mechanical seal					
18	Check seal locks are removed					
19	Check seal liquid system	see installation notes manufacturer				
20	Check equipment for soft feet					
21	Shaft alignment	only metal shims are allowed				
22	Check piping and support attached to pump	Flange deviation				
23	Vents, drain correctly installed					
24	Check earth bosses					
25	Tracing installed correctly					
26	Insulation installed correctly					
27	Check documentation	shaft alignment, foundation measurements, nameplate				
28	P&ID / PEFS check					

Construction - Presommissioning		Remarks	Signed	Signed	Signed	Date
29	Check direction of rotation	uncoupled or avoid mechanical seal runs dry				
30	Check N2 pre-charge					
31	Check barrier oil fill					
32	Check lubricant and lubricant levels					

Preservation		Remarks	Signed	Signed	Signed	Date
33	Check equipment for internal cleanliness					
34	Check proper preservation	nozzles blinded off, isolated during flushing				

Other items	Remarks
Punch item list attached	

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					



Blowers-Fans - Commissioning and Start-up checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

	Commissioning - Checks before initial run	Remarks	Signed	Signed	Signed	Date
1	Check start-up strainers					
2	Check trip and alarm settings					
3	Check instrumentation and safety devices					

	Start-up - Initial run	Remarks	Signed	Signed	Signed	Date
4	Check vibration levels					
5	Check bearing temperatures					
6	Check power consumption					
7	Check mechanical performance					
8	Check seal liquid system temperatures					
9	Check unit for leakages					
10	Remove start-up strainers					

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

References

Installation of Rotating equipment (amendments/supplements to API RP 686)

Reference to installation of rotating equipment

* foundation
* grouting
* flange alignment
* shaft alignment
* rotation check
* oil flushing
* strainers

Reference to commissioning checklist for rotating equipment in general

Centrifugal fans (amendments/supplements to ISO 13705, Annex E)

Reference to design of centrifugal fans

* dynamics
* baseplate design
* lubrication
* nameplate check

Gen : Piping - general requirements

Reference to piping requirements

* inlet piping design
* discharge piping design

Gen : Field inspection prior to commissioning of mechanical equipment

Reference to MC checklist for mechanical equipment, for each piece of equipment specific

* rotating equipment
* centrifugal compressors
* reciprocating compressors
* rotary positive displacements pumps
* blowers/fans
* centrifugal pumps
* reciprocating pumps
* rotary positive displacement pumps
* mixers

Preservation of new and old equipment standing idle

Reference to preservation of rotating equipment

* preservation oils
* cleaniness

Mixers - Construction and Pre-commissioning checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

	Construction - General checks	Remarks	Signed	Signed	Signed	Date
1	Check installation of bolts					
2	Check equipment for transport damage					
3	Check unit completeness					
4	Check whether mounting flange is correctly attached					
5	Check free movement of rotor assembly	E-motor de-energised				

	Construction - Checks after grouting	Remarks	Signed	Signed	Signed	Date
10	Check correct bolts and gaskets are used					
11	Check correct installation coupling					
12	Check coupling guard and mounting					
14	Check lubrication system					
19	Check seal					
21	Check seal liquid system	see installation notes manufacturer				
23	Shaft alignment	only metal shims are allowed				
24	Check piping and support attached to pump	Flange deviation				
25	Vents, drain correctly installed					
26	Check earth bosses					
27	Tracing installed correctly					
28	Insulation installed correctly					
29	Check documentation	shaft alignment, foundation measurements, nameplate				
30	P&ID / PEFS check					

	Construction - Presommissioning	Remarks	Signed	Signed	Signed	Date
31	Check direction of rotation	uncoupled or avoid mechanical seal runs dry (OH5)				
32	Check N2 pre-charge					
33	Check barrier oil fill					
34	Check lubricant and lubricant levels					

	Preservation	Remarks	Signed	Signed	Signed	Date
35	Check equipment for internal cleanliness					
36	Check proper preservation	nozzles blinded off, isolated during flushing				

Other items	Remarks
Punch item list attached	

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

Mixers - Commissioning and Start-up checklist

General information	
Requisition number	
Plant	
Unit	

Equipment information	Tag, manufacturer, type
Pump	
Driver	
Coupling	
Gearbox	
Mechanical seal	
Seal liquid system	
Lubrication system	

Commissioning - Checks before initial run		Remarks	Signed	Signed	Signed	Date
1	Check start-up strainers					
2	Check trip and alarm settings					
3	Check instrumentation and safety devices					

Start-up - Initial run		Remarks	Signed	Signed	Signed	Date
4	Check vibration levels					
5	Check bearing temperatures					
6	Check power consumption					
7	Check mechanical performance					
8	Check unit for leakages					

	Name	Department	Initial	Signed	Date
Engineering contractor					
Rotating Equipment engineer					
Supervisor unit/plant					

References

Gen : Installation of Rotating equipment (amendments/supplements to API RP 686)

Reference to installation of rotating equipment

* foundation
* grouting
* flange alignment
* shaft alignment
* rotation check
* oil flushing
* strainers

Reference to commissioning checklist for rotating equipment in general

Gen : Piping - general requirements

Reference to piping requirements

* inlet piping design
* discharge piping design

Gen : Field inspection prior to commissioning of mechanical equipment

Reference to MC checklist for mechanical equipment, for each piece of equipment specific

* rotating equipment
* centrifugal compressors
* reciprocating compressors
* rotary positive displacements pumps
* blowers/fans
* centrifugal pumps
* reciprocating pumps
* rotary positive displacement pumps
* mixers

Gen : Preservation of new and old equipment standing idle

Reference to preservation of rotating equipment

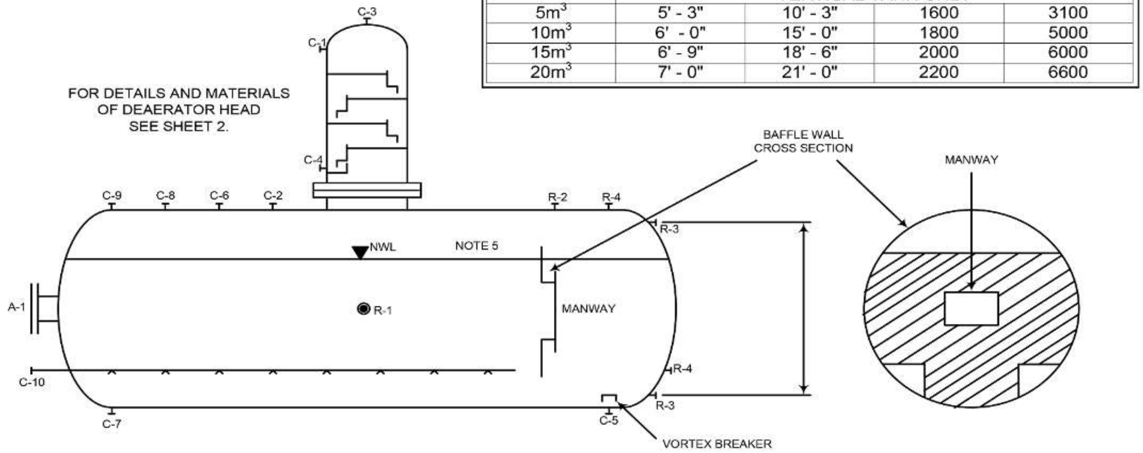
* preservation oils
* cleaniness

	DEAERATOR PROCESS DATA SHEET	Document No.	
		Sheet of	Rev.
Job		Item No.(s)\	
Item Name	**DEAERATED WATER STORAGE TANK**	No. Working	Total No. Off

NOTE: * indicates delete as necessary: ** indicates for other than code reason.

STANDARD TANK DIMENSIONS

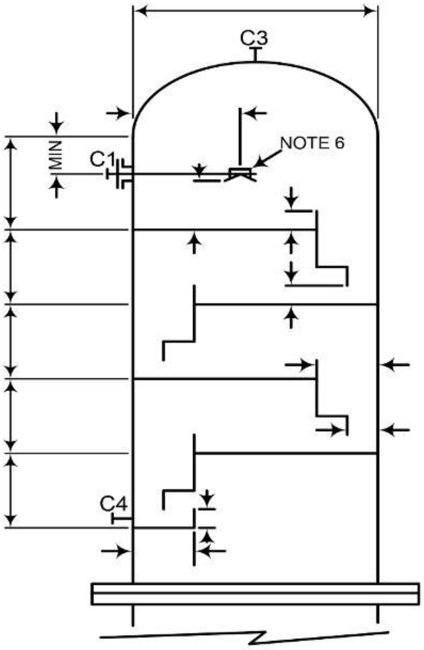
TANK	ENGLISH		METRIC	
SIZE	O.D.	LENGTH	O.D.	LENGTH
$2m^3$	VERTICAL TANK ONLY			
$5m^3$	5' - 3"	10' - 3"	1600	3100
$10m^3$	6' - 0"	15' - 0"	1800	5000
$15m^3$	6' - 9"	18' - 6"	2000	6000
$20m^3$	7' - 0"	21' - 0"	2200	6600

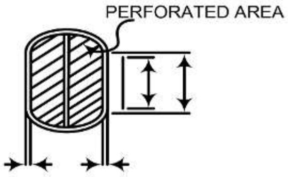


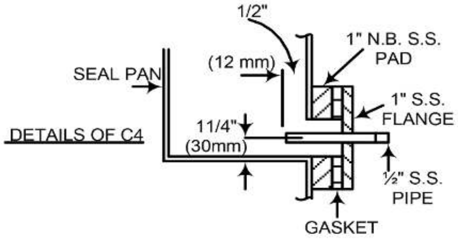
26	Shell Diameter (O.D.)					Shell Length: (T.L. - T.L.)			No. Required	
27	Center Line: *Horizontal									
28		Pressure:		Temperature:		Nozzles	Mark No.	Size In	Qty	
29		psig	bar g	°F	°C	Inlet	C-1		1	
30	Operating	3.0	0.237	223	106	Steam In	C-2		1	NOTE 4
31	Design	10.0	0.73	241	116	Vapour Out	C-3		1	
32	emergency Vacuum Design: *Yes					Sample Point	C-4		1	
33		Material		Corr. Allowance		Liquid Out	C-5		1	
34	Shell	CS		1/16" (1.6mm) }	NOTE	Cond. Out	C-6		1	NOTE 4
35	Heads	CS		1/16" (1.6MM) }	1					
36	Liner					Vent	C-			
37	Type of Heads					Drain	C-7		1	
38	Code:					Steam Out	C-			
39	Stress Relieve**: Yes					Pressure Relief	C-8		1	
40	Radiography**: No					Spill Back	C-9		1	
41	joint Efficiency:					Start Up	C-10		1	NOTE 3
42	Weight Empty:			Weight Full:						
43	Is vessel subject to mechanical vibration?				NOTE 2					
44	Insulation:	Conservation								
45	*Yes									
46	NOTES									
47	1) Corrosion Allowance on Tank Only					Manhole	A-1		1	
48	2) Water Hammer Only									
49	3) Sparge Pipe To Have Delavan Type BB Nozzles					Thermocouple	R-1	by Inst. Gp.	1	
50	and Terminate 1000mm Before Baffle.					Pressure Gauge	R-2		1	
51	4) Allow for Temp of Desuperheating Steam					Gauge Glass	R-3		2	
52	or condensate at Inlet Nozzles.					Level Control	R-4		2	Incl.LAHL/LL
53	5) Normally at 0.67 Dia. Above Base.									
54										
55						Min. Base Elevation:			Skirt Length:	
56						Material:				

	1	Date	2	Date	3	Date	4	Date	5	Date
Description										
Made/Revised by										
Checked by										
Approved Process										
proved by										

	DEAERATOR PROCESS DATA SHEET	Document No.	
		Sheet of	Rev.
Job		Item No.(s)	
Item Name	**DEAERATOR HEAD**	No. Working	Total No. Off







1. DESIGN FOR FULL VACUUM.
   OPERATING TEMPERATURE _______________
   OPERATING PRESSURE __________________
2. TRAYS TO BE SEALED TO SHELL WITH STEAM QUALITY RUBBER GASKETS.
3. C4 TO BE CAPPED FOR SITE INSTALLATION OF STANDARD SAMPLING EQUIPMENT.
4. HEAD TO BE IN S.S OR COATED M.S. BELOW 120°C MAX. EPOXY RESIN BELOW 180°C MAX. STOVED PHENOLIC RESIN (SAKAPHEN). PROCESS TO ADVICE ON SPECIFICATION & APPLICATION.
5. TRAYS TO BE IN 18/8 S/S. TRIANGULAR PERFORATION. PATTERN TO BE N.GREENING'S___ OR EQUAL.
6. SPRAY NOZZLE TO BE IN 18/8 S/S. SIMILAR TO DELAVAN WATSON S. _______________ AND WITHDRAWABLE THROUGH C-1.

NOTE: Bubbling Area Is Any Part Of Tray Within 4" Distance Of A Hole

NOZZLES	MARK	SIZE	QTY
Water Inlet	C-1		
Vent	C-3		
Sample Point	C-4	STD	1
Perf. Area/Tray			
Bubbling Area/Tray			
No. Of Trays			
Tray Thickness			
Hole Diameter			
Pitch. Approx.			
No. of Holes/Tray			

FLOWS (kg/hr)		
	NORMAL	DESIGN
Water Feed 1		
Water Feed 2		
Water Feed 3		
Heating Steam		
Start-Up Steam		
Vent Steam		
Deaerated Feed Water		

* No. Of Hole Is More Important Than Pitch

	1	Date	2	Date	3	Date	4	Date	5	Date
Description										
Made/Revised by										
Checked by										
Approved Process										
Approved by										

	ΣΩΣ A.K.C. TECHNOLOGY	**DRUM PROCESS DATA SHEET**	Doc. No:
			Item No.
		Job:	Sheet of
Distribution:	Item Name:		

NOTE: * indicates delete as necessary; ** indicates for other than Code reason

26	Shell Diameter (O.D. / I.D.):				Shell Length:			No. Required:	
27	Center+C71 Line: *Horizontal / Vertical								
28		Pressure: g		Temperature: °F / °C	Nozzles	Mark No.	Size	Number	
29	Item Number				Inlet	C-			
30	Operating								
31	Design				Vapor Out	C-			
32	Emergency Vacuum Design: * Yes / No								
33		Material	Corr. Allowance		Liquid Out:	C-			
34	Shell								
35	Heads								
36	Liner				Thermocouple	R-	by Inst. Op.		
37	Type of Heads				Pressure Gauge	R-			
38	Code:				Gauge Glass	R-			
39	Stress Relieve**:	Radiography**:			Level Control	R-			
40	Joint Efficiency:				Safety Valve	R-			
41	Density of Contents:		at	*°F / °C					
42	Weight Empty:		Weight Full:						
43	Is vessel subject to mechanical vibration? *Yes / No				Vent	C-			
44	insulation:	Type: *Frost and Personnel Protection / Cold			Drain	C-			
45	*Yes / No	/ Anticondensation / Heat Conservation			Steam Out	C-			
46	REMARKS:								
47					Manhole	A-			
48									
49									
50									
51									
52									
53					min. Base Elev'n:				
54					Material:				

Issue No.	1	Date	2	Date	3	Date	4	Date
Made/Revised by								
Checked by								
Approved - Process								
Approved by								

EFFLUENT SCHEDULE

Job No.:		Sheet of	Issue No.:	1	Date	2	Date
Job Name:			Description				
Document No.			Made/Revised by				
Project No.			Checked by				
Client			Approved- Process				
Location			Approved				
Plant							

	STREAM NUMBER	EFFLUENT SOURCE	FLUID	TOTAL FLOWRATE tonne/h.	MAIN CONTAMINANTS	CONTAMINANT kg/h. QUANTITY	CONTAMINANT CONCENTRATION	STREAM TEMP. °C.
1								
2								
3								
4								
5								
6								
7								
8								
9								
10								
11								
12								
13								
14								
15								
16								
17								
18								
19								
20								
21								
22								
23								
24								
25								
26								
27								
28								
29								
30								
31								
32								
33								
34								
35								
36								
37								
38								
39								
40								
41								
42	Notes:							
43								
44								
45								
46								
47								
48								

	PROCESS DATA SHEET	Job No. Item No.
	Job:	Sheet of

EQUILIBRIUM FLASH CALCULATION (Form 1 of 2)

Temperature: .... °F
Pressure: .... psia
Pg: .... psia

Assumed $\frac{V}{L}$ =												
	m	K	$K\frac{V}{L}+1$	$\frac{m}{K\frac{V}{L}+1}$	$K\frac{V}{L}+1$	$\frac{m}{K\frac{V}{L}+1}$	$K\frac{V}{L}+1$	$\frac{m}{K\frac{V}{L}+1}$	$K\frac{V}{L}+1$	$\frac{m}{K\frac{V}{L}+1}$	$K\frac{V}{L}+1$	$\frac{m}{K\frac{V}{L}+1}$
H_2 CO CH_4												
CO_2 C_2H_6 C_3H_8												
i-C_4H_{10} n-C_4H_{10} C_2H_4												
C_3H_6 C_6H_6												
$\sum\frac{m}{K\frac{V}{L}+1}$ =												
Calculated $\frac{V}{L}$ =												

$$\boxed{\text{Calculated}\frac{V}{L}=\frac{1}{\sum\frac{m}{K\frac{V}{L}+1}}-1}$$

Issue No.	1	Date	2	Date	3	Date	4	Date	5	Date
Made/Revised by										
Checked by										
Approved - Process										
Approved by										

	PROCESS DATA SHEET	Job No. Item No.
	Job:	Sheet of

EQUILIBRIUM FLASH CALCULATION (Form 2 of 2)

Temperature: ______ °F
Pressure: ______ psia
Pg: ______ psia

	MOLES			MOL. FRACTION	
	Feed m	Liquid L_x	Gas V_y	Liquid x	Gas y
H_2					
CO					
CH_4					
CO_2					
C_2H_6					
C_3H_8					
i-C_4H_{10}					
n-C_4H_{10}					
C_2H_4					
C_3H_6					
C_6H_6					

$$L_x = \frac{m}{K \frac{V}{L} + 1} \qquad V_y = m - L_x$$

Issue No.	1	Date	2	Date	3	Date	4	Date	5	Date
Made/Revised by										
Checked by										
Approved - Process										
Approved by										

FABRICATED EQUIPMENT SCHEDULE

Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
Job Name:			Description						
Document No.			Made/Revised by						
Project No.			Checked by						
Client			Approved- Process						
Location			Approved						
Plant									

ITEM No.	No. OFF	DESCRIPTION	DIAMETER (m)	LENGTH (m)	SURFACE (m^2/SHELL)	WEIGHT tonne/SHELL	INTERNALS/ H.E. TYPE (See Note 1)	FLUID T = Tube S = Shell	MATERIAL	DESIGN PRESSURE ($kg/cm^2.g$)	CONDITIONS TEMPERATURE (°C)	REMARKS
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				
								T				
								S				

Notes: 1. H.E. Type Coding - A.C. = Air Cooler; K = Kettle; U.B. = U - Tube Bundle; F.H. = Floating Head; F.T.S. = Fixed Tube Sheet

A.K.C. TECHNOLOGY	**FAN/COMPRESSOR PROCESS DUTY SPECIFICATION**	Document No.	
		Sheet of	Rev.
Job		Item No.(s)\	
Item Name		No. Working	Total No. Off

	OPERATING CONDITIONS PER UNIT								
1		**UNITS**							
2	Operating Case								
3	Mass Flowrate	kg/h	lb/h						
4	Standard Volumetric Flowrate	nm^3/h	std ft^3/h						
5	Volume at Suction	m^3/h	ft^3/h						
6		**INLET CONDITIONS**							
7	Pressure	bara	psia						
8	Temperature	°C	°F						
9	Molecular Weight								
10	Cp/Cv								
11	Compressibility								
12		**DISCHARGE CONDITIONS**							
13	Pressure	bara	psia						
14	Temperature	°C	°F						
15	Cp/Cv								
16	Compressibility								
17		**GAS ANALYSIS**							
18		Mol %	Mol %						
19									
20									
21									
22									
23									
24									
25									
26									
27									
28									
29									
30									
31	Entrained Liquids/Solids								
32		**PERFORMANCE**							
33	Compression Ratio								
34	Estimated Efficiency	%	%						
35	Estimated Absorbed Power	kW	hp						
36	Recommended Driver Power	kW	hp						
37		**MECHANICAL ARRANGEMENT**							
38	Compressor/Fan Type								
39	Number of Stages			**SKETCH**					
40	Gas to be kept Oil Free	Yes / No							
41									
42	Driver Type								
43									
44	Casing Design Pressure								
45	Design Pressure								
46									
47	Material - Casing								
48	- Impeller								
49									
50	Sour Service	Yes / No							
51	NOTES								
52	1. Cooling medium available:-								
53									
54									
55									

	1	Date	2	Date	3	Date	4	Date	5	Date
Description										
Made/Revised by										
Checked by										
Approved Process										
Approved by										

FRACTIONATOR CALCULATION SUMMARY

Job No:
Item No:
Sheet ____ of _____

Job:

Item Name:

Distribution

1	Operate at			psig	Units: Metric / British (delete one)							
2	Fractionation:							Reflux Ratio & No. of Plates:				
3	Ovhd Prod:		C	<	Reflux Ratio:				Actual:		; Use	
4	Bottoms:		C	<	Rectifying Plates:				Theoretical:		; Use	
5					Stripping Plates:				Theoretical:		; Use	
6	Feed					Net Bottoms						
7	Component		Mols	Mol %		Mols	Mol %	K Ksca °C psia °F	KX	Pat °C °F	Vapor Pressure	
8												
9	1											
10	2											
11	3											
12	4											
13	5											
14	6											
15	7											
16	8											
17	9											
18	10											
19												
20	Net Overheads							Ext. Rflx.	Gross Overheads			
21	Component		Mols	Mol %	K Ksca °C psia °F	KX or $\frac{Y}{K}$		Mols	Mols	Mol %	K Ksca °C psia °F	$\frac{Y}{K}$
22												
23	1											
24	2											
25	3											
26	4											
27	5											
28	6											
29	7											
30	8											
31	9											
32	10											
33												
34	Thermal Conditions of Feed:							Reboiler Vapor Quantity:				
35												
36												
37												
38												
39												
40												
41	Stream			°API	Sp.Gr. at 15°C at 60°F	Temp. °C °F	Sp.Gr. at T°	Sp. Vol. At T° cu. m/kg cu. Ft/lb		Mols/hr	Mol. Wt.	kg/hr lb/hr
42												
43												
44	Feed											
45	Net Bottoms											
46	Net Overheads											
47	Gross Overheads											
48	Gross Bottoms											
49	Reboiler Vapor											
50												
51												
52												
53	REMARKS											
54												
55												
56					Issue No.		1	Date	2	Date	3	Date
57					Made/Revised by							
58					Checked by							
59					Approved - Process							
60					Approved by							

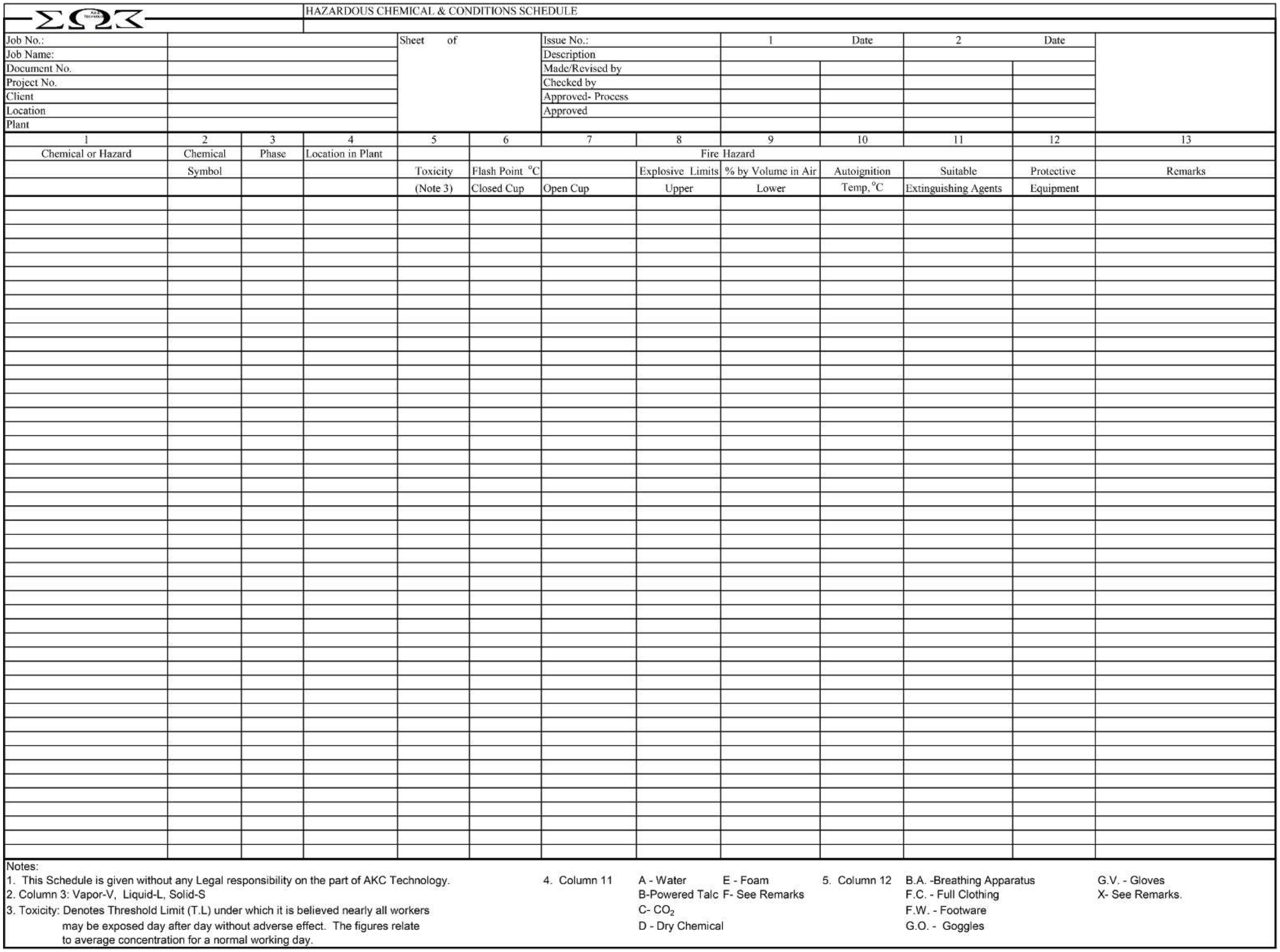
GENERAL SERVICES AND UTILITIES CHECK LIST

Job No.:		Sheet of	PREPARED BY:
Job Name:			
Document No.			DATE:
Project No.			
Client			
Location			
Plant			

SERVICE	SUSTAINED	QUANTITY PEAK	UNIT	OPERATING PRESSURE & TEMPERATURE	REMARKS
ELECTRICAL 13.8 KV			*KVA		
ELECTRICAL 2300V			*KVA		
ELECTRICAL 440V			*KVA		
MISC. LIGHTING, ST. .ETC.,] (110 V)			*KVA		
WELL WATER (CITY)			GPM		
SEA WATER			GPM		
RIVER WATER - PROCESS			GPM		
RIVER WATER - FIRE PROT.			GPM		
FUEL GAS			SCFM		
ODORIZED GAS			SCFM		
COMPRESSED AIR			SCFM		
STEAM			LB/HR		
(400 OR 475) PSIG			LB/HR		
235 PSIG			LB/HR		
150 PSIG			LB/HR		
30 PSIG			LB/HR		
CONDENSATE RETURN			LB/HR		
CONDENSATE USAGE			LB/HR		
PROCESS TRANSFER & RAW MATERIALS					
CAUSTIC (- %)			LB/HR		
BRINE (-%)			LB/HR		
LPG			LB/HR		
SITE DEV. (AREA FILL)					
RAILROADS					
BLOCK ACCESS ROADS					
SANITARY SEWERS			GPM		
			GPM		
WASTE WATER					

*ASSUME 1 HORSE POWER = 1KVA

General services and utilities checklist.



HAZARDOUS CHEMICAL & CONDITIONS SCHEDULE

Job No.:		Sheet of	Issue No.:	1	Date	2	Date
Job Name:			Description				
Document No.			Made/Revised by				
Project No.			Checked by				
Client			Approved- Process				
Location			Approved				
Plant							

1	2	3	4	5	6	7	8	9	10	11	12	13
Chemical or Hazard	Chemical	Phase	Location in Plant		Fire Hazard							
	Symbol			Toxicity	Flash Point °C		Explosive Limits	% by Volume in Air	Autoignition	Suitable	Protective	Remarks
				(Note 3)	Closed Cup	Open Cup	Upper	Lower	Temp, °C	Extinguishing Agents	Equipment	

Notes:

1. This Schedule is given without any Legal responsibility on the part of AKC Technology.
2. Column 3: Vapor-V, Liquid-L, Solid-S
3. Toxicity: Denotes Threshold Limit (T.L) under which it is believed nearly all workers may be exposed day after day without adverse effect. The figures relate to average concentration for a normal working day.
4. Column 11
   - A - Water
   - B-Powered Talc
   - C- CO_2
   - D - Dry Chemical
   - E - Foam
   - F- See Remarks
5. Column 12
   - B.A. -Breathing Apparatus
   - F.C. - Full Clothing
   - F.W. - Footware
   - G.O. - Goggles
   - G.V. - Gloves
   - X- See Remarks.

Hazardous chemical and conditions schedule.

HEAT & MASS BALANCES

Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
			Made/Revised by						
Job Name:			Checked by						
			Approved- Process						
			Approved by						

	* delete unit as necessary Stream ↓	Units	→	°API	K	M.W.	FLOWING Sp.Gr.	* °F °C	* barg psig	B.P.S.D. at 60°F	FLOWING GPM	Mol. %	Mol/h	* kg/h lb/h	* kcal/kg Btu/lb	* 106 kcal/h 106 Btu/h
1																
2																
3																
4																
5																
6																
7																
8																
9																
10																
11																
12																
13																
14																
15																
16																
17																
18																
19																
20																
21																
22																
23																
24																
25																
26																
27																
28																
29																
30																
31																
32																
33																
34																
35																
36																
37																
38																
39																
40																
41																
42																
43																
44																
45																
46																
47																

A.K.C. TECHNOLOGY

HEAT EXCHANGER RATING SHEET

PART 1 - TO BE FILLED IN COMPLETELY

Document No.

Sheet of Rev.

Items No. (s)

1 Exchanger Name: *

2 Job: * * Location: * Arrangement: Horiz = 0; Vert = 1

3 Type of Unit: F.T.S.=0; FLTG.HEAD=1; U-TUBE=2; FOR KETTLES ADD 3 Units: S.I. 0; Engineering = 1

4 Hot Side: Either=0; Shell=1; Tubes=2

5	Units	S.I.	Engineering	HOT FLUID	COLD FLUID
6	Fluid Circulated	-	-		
7	Total Fluid Entering (Normal)	kg/hr	lb/hr		
8	Flow Margin	%	%		
9	Temperature In/Out	°C	°F		
10	Max Pressure Drop at line 8	bar	psi		
11	Inlet Pressure Operating/Design	barA	psia		
12	Normal Heat Load	kW	Btu/hr		
13	Heat Load Margin	%	%		
14	Design Temperature	°C	°F		
15	Corrosion Allowance	mm	ins		
16	Fouling Resistance	°Cm^2/W	°F hr ft^2/Btu		
17	Line N.B. In/Out	mm	ins		
18	Part 2				
19	GAS (AD VAPOR) IN	-	-	■	■
20	Flow of Vapor and Gas at Inlet	kg/hr	lb/hr		
21	Code Number	-	-		
22	Molecular Weight	-	-		
23	Thermal Conductivity	W/m°C	Btu/hr ft°F		
24	Specific Heat	kJ/kg°C	Btu/lb°F		
25	Compressibility Factor	-	-		
26	Viscosity	cP	lb/ft hr		
27	LIQUID IN	-	-	■	■
28	Total Flow of Liquid at Inlet	kg/hr	lb/hr		
29	Code Number	-	-		
30	Thermal Conductivity	W/m°C	Btu/hr ft °F		
31	Specific Heat	kJ/kg°C	Btu/lb °F		
32	Density	kg/m^3	lb/ft^3		
33	Viscosity	cP	lb/ft hr		
34	Viscosity at Ave. Temp. other side	cP	lb/ft hr		
35	CONDENSATION AND VAPORIZATION	-	-	■	■
36	Fluid Condensed or Vaporized	kg/hr	lb/hr		
37	Code Number	-	-		
38	Molecular Weight	-	-		
39	Thermal Conductivity	W/m°C	Btu/hr ft °F		
40	Specific Heat	kJ/kg°C	Btu/lb °F		
41	Density (Hot Fluid)	kg/m^3	lb/ft^3		■
42	Compressibility Factor (Cold Fluid)	-	-		
43	Viscosity	cP	lb/ft hr		
44	Latent Heat	kJ/kg	Btu/lb		
45	Surface Tension	N/m	lb/ft	■	
46	Expansion Coefficient	1/°C	1/°F	■	
47	% Condensate Forming Film	%	%		■
48	Thermal Conductivity	W/m°C	Btu/hr ft °F		■
49	Density	kg/m^3	lb/ft^3		■
50	Viscosity	cP	lb/ft hr		
51	Temps for Phase Change Start/Finish	°C	°F		
52	LMTD weighting Factor 'F'	-	-		

53	PART 3												
54	Number of Points in Table Below												
55	% Heat Load	%	%	1	2	3	4	5	6	7	8	9	10
56	Hot Fluid Temperature	°C	°F										
57	Cold Fluid Temperature	°C	°F										

58 PART 4 - CODE NUMBERS TO BE FILLED IN COMPLETELY

59	Material for	Description	Code No.	O.D.	Thickness	Length	Passes	Number	Contents Lethal?
60	Tubes								Yes/No
61	Shell							■	Yes/No
62	Floating Head					Tube Pitch = mm/in.	o	Δ	□
63	Channel								
64	Tube Sheets								

NOTES

Issue No.	1 Date	2 Date	3 Date	4 Date	5 Date	
Made/Revised by						
Checked by						
Approved- Process						
Approved by						

	HYDROCARBON DEWPOINT CALCULATION	Job No: Item No.
	Job:	Page of

Pressure: .... psia

Heaviest Compound: PROPANE

Convergence Pressure: .... psia

	y	Tb	LIGHT COMPONENT		HEAVY COMPONENT		ASSUMED TEMPERATURE								
							°F		°F		°F		°F		°F
			FeL	TbFeL	FeH	TbFeH	K	y/K	K	y/K	K	y/K	K	y/K	x
H_2		67	1.000	67.000											
N_2		130	0.0229	2.972	0.105	13.61									
CO		172	0.0046	0.797	0.180	31.00									
CH_4		201	0.0019	0.383	0.244	49.02									
C_2H_4		305			0.548	167.0									
CO_2		330			0.638	210.6									
C_2H_6		332			0.646	214.3									
C_2H_2		340			0.676	229.9									
C_3H_6		406			0.954	387.3									
C_3H_8		416			1.000	416.0									
		1.000			Σ y/K										1.000
					x for H_2										
					$\Sigma (yFe_L/K)$										
					$\Sigma (yFe_H/K)$										
					E.B.PL										
					E.B.PH										
					P_g										

Estimate of Dewpoint: .... °F

at Estimate Pg of: .... psia

Issue No.	1	Date	2	Date	3	Date
Made/Revised by						
Checked by						
Approved - Process						
Approved by						

LINE LIST SCHEDULE

Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date	4	Date	5	Date	6	Date
Job Name:			Description												
Document No.			Made/Revised by												
Project No.			Checked by												
Client			Approved- Process												
Location			Approved												
Plant															

1	2	2A	3	4	5		6		7	8		9		10		11					12	13	14	14a	15	16
ISS	LINE		NOM.	PIPING	FLUID		LINE	ROUTING	PLANT	OPERATING		DESIGN		INSULATION			FIELD	TEST			FIELD	CHEMICAL	REMARKS	STRESS	ELD	ISO. DRG.
COLUMN	No.		PIPE	SPEC.	DENSITY	PHASE	FROM	TO	CONDITION	PRESS.	TEMP.	PRESS.	TEMP.	CLASS	NOM. THICK	TEST					TEST	CLEANING		CATEGORY	DRG.	No.
DATA			SIZE													CLASS	CLASS	PRESS.	CLASS	PRESS.	LIMITED BY				No.	
CHANGES																										

UNITS: PRESSURE bar g p.s.l.g kg/cm^2 g TEMPERATURE: °C °F NOMINAL PIPE SIZE: Inches mm Insulation Thickness: Inches mm.

Column

1 Enter Issue Number in small box. Indicate in outside box number(s) of column(s) containing revision

2A Blocking this column indicates that line is Process Critical: I.e. requires routing feedback to Pressure.

5 Indicate V for vapors: L for liquids: L/V for mixed phases or make special reference to other conditions (e.g. slurries, etc.)

7 NO = Normal Operation, defined as the coincident pressure and temperature conditions prevailing for the majority of the plant life (insulation to be related to the temperature stated here) MO = Maximum operation, defined as the maximum sustained conditions (material selection is usually related to the temperature stated here).
SV = Relief Valve Set Pressure and coincident temperature
(minimum basis for "Design P and T") PSO = Pump Shut-Off head (at reduced or no flow) where no pressure relief is provided XT ( ) = Transient conditions which can occur for the total period of time shown in paranthesis, in hours, during an operational year (abbreviated to T_{10} or T_{50}, as appropriate. If Job Piping Code incorporates ANSI B31.3)
RT = Relief Valve Blowing conditions, taken as allowing maximum pressure / stress relaxation permitted by Job Piping Code.

8 Coincident pressures and temperatures corresponding to the process cases quoted under Column 7 (the highest temperature) quoted under Column 8 is used for calculating thermal stresses.

9 The pressure and temperature upon which the mechanical design of the line will be based (used for pipewall thickness, flange, gasket and bolting design).

10 Insulation class and thickness to contract requirements:
Classs 0 No insulation
1 Heat Conservation
2 Special Heat Conservation (e.g. controlled heat loss for maintaining temperatures).

3 Personnel Protection Insulation
4 Personnel Protection Guards only (I.e. Insulation not permissible).
5 Cold Conservation
6 Special Cold Conservation (e.g. anti-condensation only or controlled heat gain).
7 Frost Protection
8 Acoustic Lining
9 Acoustic Lagging
11 To indicate Jacketed Line
12 To indicate Steam Traced Line
14 To indicate Electrically Traced Line
Example: Class 2/12 denotes special heat conservation insulation for a steam traced line.

11 Field testing to Contract Specification No:

13 Cleaning of Pipework (in addition to standard pre-commissioning clean-out). For co-ordination
Type 1 Manual (wire brush or equivalent)
2 Pigging (including wire brush pig)
3 Sandblast or Shotblast
4 Pickling - Shop
5 Pickling - Site
6 Chemical Cleaning (by circulation)
7 Steam Blowing
8 By Equipment Vendor

14A Enter the Stress Category as advised by Piping Streee Engineer.

15 For latest issue of listed drawing numbers see Drawing Number Register.

16 For latest issue of listed drawing numbers see Drawing Number Register.

LINE SCHEDULE
TITLE:

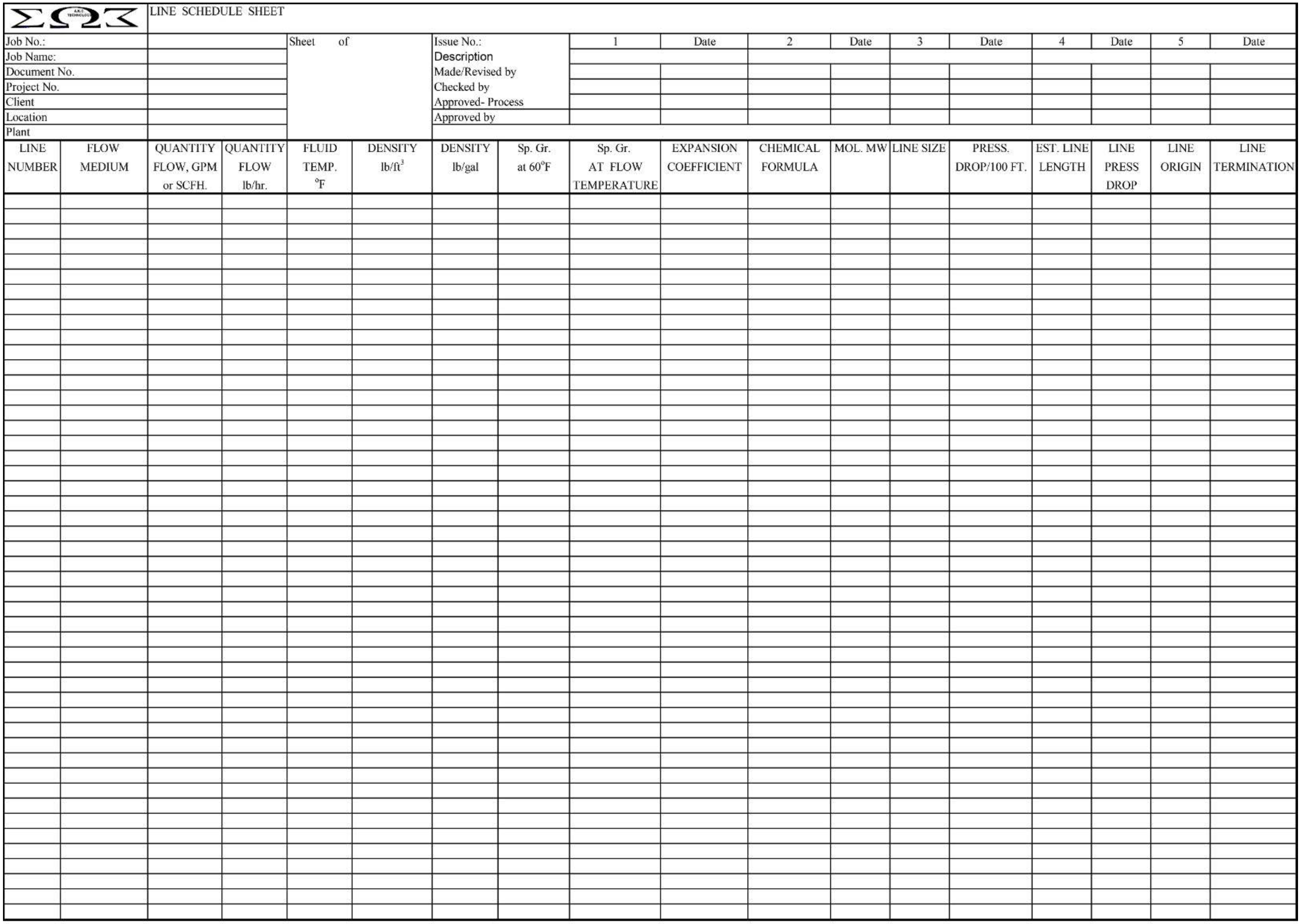
Job No.:
Job Name:
Document No.
Project No.
Client
Location
Plant

Sheet of

Issue No.:	1	Date	2	Date	3	Date
Description						
Made/Revised by						
Checked by						
Approved- Process						
Approved by						

LINE NUMBER	SIZE INCHES	FROM	TO	LINE CLASS OR CODE	INSULATION CODE	TEST PRESSURE, PSIG.	SPECIAL REMARKS

Line Schedule.



LINE SCHEDULE SHEET

Job No.:
Job Name:
Document No.
Project No.
Client
Location
Plant

Sheet of

Issue No.:	1	Date	2	Date	3	Date	4	Date	5	Date
Description										
Made/Revised by										
Checked by										
Approved- Process										
Approved by										

LINE NUMBER	FLOW MEDIUM	QUANTITY FLOW, GPM or SCFH.	QUANTITY FLOW lb/hr.	FLUID TEMP. °F	DENSITY lb/ft^3	DENSITY lb/gal	Sp. Gr. at 60°F	Sp. Gr. AT FLOW TEMPERATURE	EXPANSION COEFFICIENT	CHEMICAL FORMULA	MOL. MW	LINE SIZE	PRESS. DROP/100 FT.	EST. LINE LENGTH	LINE PRESS DROP	LINE ORIGIN	LINE TERMINATION

Line schedule sheet.

LINE SUMMARY TABLE

TITLE:

Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
Job Name:			Description						
Document No.			Made/Revised by						
Project No.			Checked by						
Client			Approved- Process						
Location			Approved						
Plant									

NUMBER	MATERIAL	FROM	TO	LINE SIZE	OPERATING TEMP. °F	OPERATING PRESS. PSIG	TEST PRESS. PSIG	QUANTITY				VEL. F.P.S.	PRESSURE DROP		REMARKS	NO.
								LB/HR.	G.P.M. @60 oF	GPM (HOT)	C.F.S (HOT)		PER 100 ft. PSI	TOTAL PSI		

Line summary table

MASS BALANCE

Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
Job Name:			Description						
Document No.			Made/Revised by						
Project No.			Checked by						
Client			Approved- Process						
Location			Approved						
Plant									

	STREAM NUMBER		▽		▽		▽		▽		▽		▽		▽	
	STREAM NAME															
	PHASE															
	TEMPERATURE (°C)															
	COMPONENT	MOL. WT.	kg. mol/hr	Mol. %	kg. mol/hr	Mol. %	kg. mol/hr.	Mol. %	kg. mol/hr.	Mol. %	kg. mol/hr.	Mol. %	kg. mol/hr.	Mol. %	kg. mol/hr.	Mol. %
1	Water															
2	Hydrogen															
3	Nitrogen															
4	Carbon monoxide															
5	Carbon dioxide															
6	Methane															
7	Acetylene															
8	Ethylene															
9	Ethane															
10	Propyne															
11	Propadiene															
12	Propylene															
13	Propane															
14	Isobutane															
15	Isobuene															
16	n-Butane															
17	1-Butene															
18	1,3-Butadiene															
19	1-Pentene															
20	neo-Pentane															
21	Isopentane															
22	n-Pentane															
23	Toluene															
24	m-Xylene															
25	Methanol															
26	MTBE															
27	TBA															
28	DIISO															
29	TOTAL (kg mol/hr)															
30																
31																
32																
33																
34																
35																
36																
37																
38																
39																
40																
41																
42	Notes:															
43																
44																
45																
46																
47																
48																

MECHANICAL EQUIPMENT SCHEDULE

Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
Job Name:			Description						
Document No.			Made/Revised by						
Project No.			Checked by						
Client			Approved- Process						
Location			Approved						
Plant									

ITEM No.	No. OFF	SERVICE DESCRIPTION	FLUID	S.G/M.W	CAPACITY (m^3/h) (Turbine - Te/hr)	PRESSURE $(Kg/cm^2\ g)$		INLET TEMP. (°C)	NPSHA (m)	TYPE (see Note 1)	MATERIALS	ABSOLUTE POWER (kW)	SPEED (RPM)/ REMARKS
						IN	OUT						

Notes: 1. Rotary Equipment Type Coding: - A = Axial; C = Centrifugal; M = Metering; R= Reciprocating; S = Screw

2. All drives electric motor unless specified otherwise.

PIPE LINE LIST
TITLE:

Job No.:		Sheet of		Issue No.:	1	Date	2	Date	3	Date
Job Name:				Description						
Document No.				Made/Revised by						
Project No.				Checked by						
Client				Approved- Process						
Location				Approved						
Plant										

LINE NUMBER	FROM	TO	TEMPERATURE (oF) OPERATING	DESIGN	PRESSURE (PSIG) OPERATING	DESIGN	VELOCITY FT/SEC.	PRESSURE/100 FT. PD/100 FT.	FLOW$^{(1)}$	P & ID NUMBER

NOTES: 1. FLOW UNITS : G= gpm; C= acfm; P = lbs/hr.

Pipe line list

PIPE LIST
TITLE:

Job No.:		Sheet of		Issue No.:	1	Date	2	Date	3	Date
Job Name:				Description						
Document No.				Made/Revised by						
Project No.				Checked by						
Client				Approved- Process						
Location				Approved						
Plant										

LINE NUMBER	FROM	TO	TEMPERATURE (oF)		PRESSURE (PSIG)		VELOCITY FT/SEC.	PRESSURE/100 FT.	FLOW(1)	P & ID NUMBER
			OPERATING	DESIGN	OPERATING	DESIGN		PD/100 FT.		

NOTES: 1. FLOW UNITS : G= gpm; C= acfm; P = lbs/hr.

PIPING PROCESS CONDITIONS SUMMARY

Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
Job Name:			Description						
Document No.			Made/Revised by						
Project No.			Checked by						
Client			Approved- Process						
Location			Approved						
Plant									

INDEX NUMBER	MATERIAL AND COMPOSITION	FLUID SYMBOL	HAZARDOUS MATERIAL CLASSIFICATION	FLUID STATE LIQUID/GAS	MAXIMUM PIPE SIZE INCHES	MAXIMUM OPERATING TEMPERATURE, °F	CONTINUOUS RANGE PRESSURE, PSIA	PIPE SPECIFICATION	NOTES

PLATE HEAT EXCHANGER DATA SHEET

	Document No.	
	Sheet of	Rev.
Job	Item No.(s)	
Item Name	No. Working	Total No. Off

1	PROCESS DESIGN	UNITS		FLUID 1	FLUID 2	FLUID 3
2	Fluid Circulated	-	-			
3	Total Fluid Entering (Normal)	kg/h	lb/h			
4	Flow Margin	%	%			
5	Inlet Vapor and Gas (&MW)	kg/h	lb/h			
6	Inlet Liquid	kg/h	lb/h			
7	Fluid Vaporized/Condensed (&MW)	kg/h	lb/h			
8	Temperature In/Out	°C	°F			
9	Max. Pressure Drop	bar	psi			
10	Inlet Pressure (Operating)	bar g	psig			
11	Normal Heat Load	kW	Btu/h			
12	Heat Load Margin	%	%			
13	Fouling Resistance	m^2 °C/W	ft^2°Fh/Btu			
14	FLUID PROPERTIES					
15	Specific Heat	kJ/kg°C	Btu/lb°F			
16	Thermal Conductivity	W/m°C	Btu/hr ft°F			
17	Density	kg/m^3	lb/ft^3			
18	Viscosity	cP	lb/hr.ft			
19	Liquid Viscosity (at Temp.)	cP	lb/hr.ft			
20	Latent Heat	kJ/kg	Btu/lb			
21	ENGINEERING DESIGN					
22	Process Design Pressure	bar g	psig			
23	Process Design Temperature	°C	°F			
24	Corrosion Allowance (Header)	mm	in.			
25	Line N.B. In/Out	in	in.			
26	Minimum Flow Passage	mm	in.			
27	Plate Material					
28	Frame Material					
29	Gasket Material					
30	Desgin Code					

31	NOTES
32	
33	1. Unless otherwise stated fluid properties are for mean fluid temperatures.
34	2. Sour Service Yes / No
35	
36	
37	
38	
39	
40	
41	
42	
43	
44	
45	
46	
47	
48	
49	
50	
51	
52	
53	
54	

	1	Date	2	Date	3	Date
Description						
Made/Revised by						
Checked by						
Approved Process						
Approved by						

Made Checked Date	A.K.C. TECHNOLOGY	**PROCESS DATA SHEET**	Job No. Item No.
		Job	Sheet of

CALCULATION OF PRESSURE DROP IN FIXED CATALYST BEDS

REQUIRED DATA

VESSEL NO:

SERVICE:

CATALYST:	VOLUME REQUIRED	V		ft³
	MANUFACTURER & NO.			
	DIMENSIONS			
	EQUIV. PARTICLE DIA.	D_p		ft.
	SHAPE/SIZE FACTOR	S_f		
GAS DATA:	FLOW RATE	W		lb/h
		W^2		$(lb/h)^2$
	MIN. MOLECULAR WEIGHT			
	BED OUTLET PRESSURE	P		psia
	MAX. BED TEMPERATURE			°C
	ABS. TEMPERATURE (°K = °C + 273)	T		K

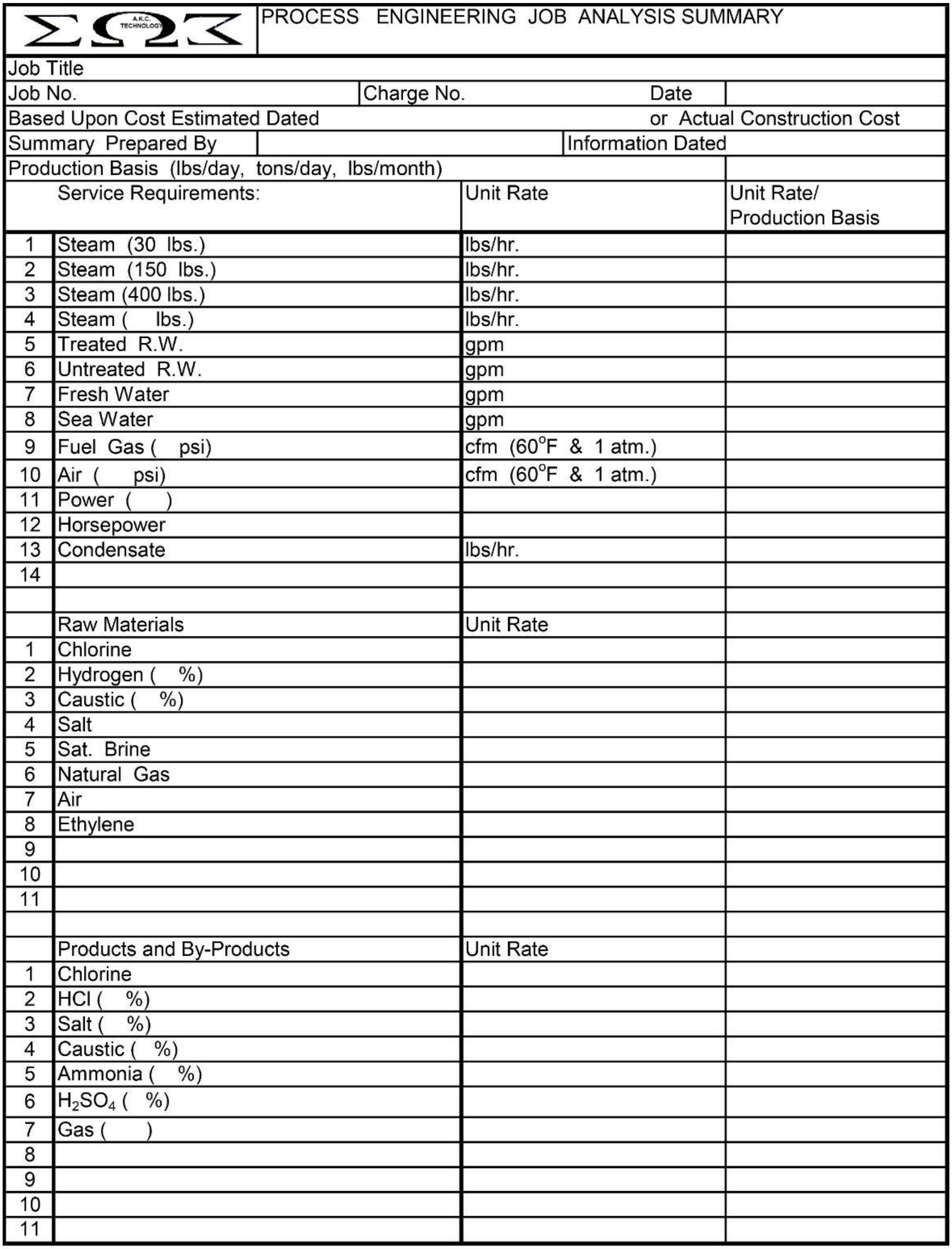
CALCULATION

$$k_1 = \frac{(T)(S_f)}{(M)(P)} = \underline{\hspace{6cm}} =$$

	SYMBOLS	UNITS	CALCULATION	CASE 1	CASE 2	CASE 3	CASE 4
BED DIAMETER	D	ft					
BED C.S.A.	A	ft^2	$0.7854D^2$				
BED DEPTH	L	ft.	V/A				
SUPERFICIAL MASS FLOW	G	$lb/h.ft^2$	W/A				
MASS FLOW FACTOR	G_f	-					
D_pG		-					
REYNOLDS NO. FACTOR	Re_f	-					
k_2	k_2	-	k_1Re_f				
CLEAN PRESSURE DROP/UNIT DEPTH	ΔP_c	psi/ft.	k_2Gf				
CLEAN OVERALL PRESSURE DROP	ΔP_c	psi	$L\Delta Pc$				
SAFETY/FOULING FACTOR	f	-		1.25/			
FLOW SHEET PRESSURE DROP	ΔP	psi	$f\Delta Pc$				

* IF VALUE OF DpG > 200 (750 FOR RING CATALYSTS) PUT Ref = 1.0
IF VALUE OF DpG < 200 (750 FOR RING CATALYSTS) CALCULATE Re = DpG/μ
(WHERE μ = DYNAMIC VISCOSITY OF GAS, lb/ft.hr)

	1	2	3	4	5	6
Description						
Made/Revised by						
Checked by						
Approved Process						
Approved by						



A.K.C. TECHNOLOGY

PROCESS ENGINEERING JOB ANALYSIS SUMMARY

Job Title

Job No. | Charge No. | Date

Based Upon Cost Estimated Dated or Actual Construction Cost

Summary Prepared By | Information Dated

Production Basis (lbs/day, tons/day, lbs/month)

	Service Requirements:	Unit Rate	Unit Rate/ Production Basis
1	Steam (30 lbs.)	lbs/hr.	
2	Steam (150 lbs.)	lbs/hr.	
3	Steam (400 lbs.)	lbs/hr.	
4	Steam (lbs.)	lbs/hr.	
5	Treated R.W.	gpm	
6	Untreated R.W.	gpm	
7	Fresh Water	gpm	
8	Sea Water	gpm	
9	Fuel Gas (psi)	cfm (60°F & 1 atm.)	
10	Air (psi)	cfm (60°F & 1 atm.)	
11	Power ( )		
12	Horsepower		
13	Condensate	lbs/hr.	
14			
	Raw Materials	Unit Rate	
1	Chlorine		
2	Hydrogen (%)		
3	Caustic (%)		
4	Salt		
5	Sat. Brine		
6	Natural Gas		
7	Air		
8	Ethylene		
9			
10			
11			
	Products and By-Products	Unit Rate	
1	Chlorine		
2	HCl (%)		
3	Salt (%)		
4	Caustic (%)		
5	Ammonia (%)		
6	H_2SO_4 (%)		
7	Gas ( )		
8			
9			
10			
11			

Process Engineering job analysis summary.

	PUMP CALCULATION SHEET	Document No.	
		Sheet of	Rev.
Job		Item No. (s)	
Item Name.		No. Working	Total No. off

		UNITS		CASE I	CASE II	SKETCH OF PUMP HOOK-UP
1						
2	Liquid Pumped					
3	Corrosion/Erosion					
4	Due To					
5	Operating Temp. (T)	°C	°F			
6	Specific Gravity at T					
7	Viscosity	cP	cP			
8	Vapor Pressure at T	bar a	psi a			
9	Normal mass Flowrate	kg/h	lb/h			
10	Normal Vol. Flowrate	m^3/h	gpm			
11	Min. Vol. Flowrate					
12	Design Vol. Flowrate	m^3/h	gpm			
13	SUCTION CONDITION					
14	Pressure at Equipment	barg	psi g	+	+	
15	Static Head	bar	psi	+ / -		
16	Total - Lines 14 + 15	bar	psi	+	+	
17	Suction Line ΔP	bar	psi	-	-	
18	Filter/Strainer ΔP	bar	psi	-	-	
19						
20	Total Suction Pressure	bar g	psi g	+	+	
21	DISCHARGE CONDITION					
22	Pressure at Equipment	bar g	psi g	+	+	
23	Static Head	bar	psi	+ / -	+ / -	
24						
25	Exchanger ΔP	bar	psi	+	+	
26						
27	Furnace ΔP	bar	psi	+	+	
28	Orifice ΔP	bar	psi	+	+	
29	Control Valve ΔP	bar	psi	+	+	
30						
31	Line ΔP	bar	psi	+	+	
32						
33	Total Discharge Press.	bar g	psi g	+	+	
34	Differential Pressure	bar	psi			
35	Differential Head	bar	psi			
36	NPSH					NOTES
37	Total Suction Pressure	bar a	psi a			1. Pump shut-ff head not to exceed………
38	Vapor Pressure	bar a	psi a			
39	NPSH - Lines 37 - 38	bar a	psi a			2. Relief valve on pump discharge to be set at ……………
40	=	m	ft.			
41	Safety Margin	m	ft.			3. Pump case design pressure…………
42	NPSH - Lines 40-41	m	ft.			design temperature……..
43	Hydraulic Power	kW	Hp			
44	Estimated Efficiency	%	%			4. Sealing/flushing fluid available:
45	Estimated Abs. Power	kW	Hp			
46	Type of Pump					5. Cooling medium available:
47	Drive					
48						6. Insulation required:
49	Material - Casing					
50	- Impeller					7. Start-up/commissioning fluid SG.
51	-Shaft					
52						
53	Sour Service	Yes/No				
54	HEAD	m = 10.2 x bar /SG	m = 10 x kg/cm2 / SG	ft = 2.31 x psi / SG		
55	VOLUME	m3/h x SG x 1000 = kg/h	igpm x SG x 600 = lb/h			
56	POWER	kW=m^3/h x bar/36.0	kW=m^3/h x kg/cm^2 /36.71			Hp = igpm x psi/1427

57		1	Date	2	Date	3	Date	4	Date	5	Date	
58	Description											
59	Made/Revised by											
60	Checked by											
61	Approved Process											
62	Approved											

PUMP SCHEDULE

Job No.:		Sheet of	
Job Name:			
Document No.			
Project No.			
Client			
Location			
Plant			

Issue No.:	1	Date	2	Date	3	Date
Description						
Made/Revised by						
Checked by						
Approved- Process						
Approved						

1	Item Number.									
2	Item Name									
3	Number Off: No. Running /Total Installed									
4	Preferred Type	Units								
5	Liquid Pumped									
6	Operating - Temperature	°C	°F							
7	Specific Gravity at Operationg Temperature									
8	Viscosity at Operating Temperature	cP	lb/ft.h							
9	Vapor Pressure at Operating Temperature	bara	psia							
10	Design Capacity (per pump)	m^3/h	ft^3/h							
11	Normal Capacity (per pump)	m^3/h	ft^3/h							
12	Discharge Pressure	bar g	psig							
13	Suction Pressure	bar g	psig							
14	Differential Head	bar /m	psi/ft							
15	Minimum NPSH Available / Required	m	ft.							
16	Hydraulic Power (per pump)	kW	hp							
17	R.P.M	-	-							
18	Estimated Pump Efficieny	-	-							
19	Estimated Shaft Power (per pump)	kW	hp							
20	Type of Driver	-	-							
21	Driver Rated Power (per pump)	kW	hp							
22	Design Power Consumption (per pump)	kW	hp							
23	Normal Power Consumption (per pump)	kW	hp							
24	Max. Shut-Off Head	bar	psi							
25										
26	Material: Casing									
27	Impeller									
28	Shaft									
29	Sour Service									
30	Packing Type / Mechanical Seal Type									
31										
32	Services / Utilities									
33	Cooling Water									
34	Seal / Flush Fluid	m^3/h	ft^3/h							
35	Quench	m^3/h	ft^3/h							
36	Notes									
37										
38										
39										
40										
41										
42										
43										
44										
45										

A.K.C. TECHNOLOGY

RELIEF DEVICE PHILOSOPHY SHEET

DOCUMENT / ITEM REFERENCE

EQUIPMENT No.:

DATE:

SHEET No.: OF

CHECKED BY:

MADE BY:

DESIGN CODES: VESSELS EXCHANGERS LINES

OTHER REQUIREMENTS

BASIS FOR CALCULATION:

SET PRESSURE, PSIG: MAX. BACK PRESSURE: (a) BEFORE RELIEVING (b) WHILE RELIEVING

NORMAL CONDITIONS UNDER RELIEF DEVICE:			Calculated	
STATE: TEMPERATURE, °F: PRESSURE, PSIG:	POSSIBLE CAUSE?	FLUID RELIEVED	RELIEF RATE, lb/h	ORIFICE AREA, in^2.
HAZARDS CONSIDERED				
1. Outlets blocked				
2. Control Valve malfunction				
3. Machine trip/ overspeed/density change				
4. Exchanger tube rupture				
5. Power failure/ Voltage dip				
6. Instrument air failure				
7. Cooling failure				
8. Reflux failure				
9. Abnormal entry of volatile liquid				
10 . Loss of liquid level				
11. Abnormal chemical reaction				
12. Boxed in thermal expansion				
13. External fire				
14. (specify)				
15. (specify)				
16. (specifty)				

SELECTED DESIGN CASE:

RELIEVED FLUID: STATE DENSITY / MW: TEMPERATURE: Cp/Cv:

COMPOSITION: :FLASHING

RELIEF RATE REQUIRED, lb/h: ORIFICE SELECTED: AREA, in^2: TYPE:

ACTUAL CAPACITY, lb/h

REMARKS/SKETCH

Issue No:	1	Date	2	Date	3	Date	
Description							
Made/Revised by							
Checked by							
Approved- Process							
Approved							

A.K.C. TECHNOLOGY

TANK & VESSEL AGITATOR DATA SHEET

Project Name:

Drawing No:
Project No.:
sheet …….. Of ……..

Equipment No:
No. Off:
Associated Vessel/Tank* Item No:

	Description	Made/Revised by	Checked by	Approved Process	Approved
1					
Date					
2					
Date					
3					
Date					

PROCESS DEPT. INFORMATION

PROCESS

1. Largest and smallest charge:
2. Components added during mixing:
3. Agitator operating while vessel is being filled or product withdrawn? Yes / No*
4. If continuous, throughput per hour:
5. Process duty: Mixing liquids / Dissolving / Suspensions / Emulsions / Gas absorption / Homogenisation*
6. Mixing effect: Violent / Medium / Moderate* Time available for mixing:
7. Working Pressure: Design Pressure:
8. Working Temperature: Design Temperature:
9. Special Remarks:

MIXING

10. Components % by weight:
11. Temperature during mixing:
12. Specific gravity of components:
13. Viscosity of mixing at mixing temperature:
14. Specific gravity of product at mixing temperature:
15. Size of solid particles:
16. Special Remarks:

DESIGN DEPT. INFORMATION

IMPELLER

17. Type of impeller:
18. Number of impellers on shaft:
19. Position of impellers:
20. Distance between shaft end & vessel: Bottom bearing:
21. Preferred impeller speed: Shaft diameter:
22. Type of drive: Direct / Vee belt / Fluid*
23. Type of seal: Vapor / Packed gland / Mechanical / Easy replacement*
24. Method of installation: Assembled in / Assembled out* of vessel
25. Entry position:
26. Materials of construction:
27. Absorbed HP / KW*: Installed HP / KW*:
28. Other information:

TANK OR VESSEL

29. Dimensions: Capacity:
30. Coils, baffles, etc.:
31. Fixing agitator (beams, flanges, etc.):
32. If at atmospheric pressure: Closed / Open* Can stuffing box be greased? Yes / No*
33. Headroom available above agitator:
34. Other information:

ELECT.

35. Agitator installed: Indoors/Outdoors Motor enclosure:
36. Electrical specification:
37. Any other electrical information:

GENERAL

38. Motor to be included: Yes / No*
39. Motor Will / Will not* be sent to manufacturer for assembly and allignment.
40. Threads: Unified /*
41. All rotating parts must be strictly guarded to BS 1649/ASME
42. Fixing bolts supplied by:
43. Net weight including motor:
44. Witnessed run in air:

45. General notes:

NOTE * indicates delete as necessary.

A.K.C. TECHNOLOGY

TANK PROCESS DATA SHEET

Job:

Doc. No:
Item No.
Sheet of

Item Name:

NOTE: * indicates delete as necessary; ** indicates for other than Code reason

5										
10										
15										
20										
25										
26	Shell Diameter (O.D. / I.D.):					Shell Length:			No. Required:	
27	Center Line: *Horizontal / Vertical									
28		Pressure:	g	Temperature:	°F / °C	Nozzles	Mark No.	Size	Number	
29	Item Number					Inlet	C-			
30	Operating									
31	Design					Vapor Out	C-			
32	Emergency Vacuum Design: * Yes / No									
33		Material		Corr. Allowance		Liquid Out:	C-			
34	Shell									
35	Heads									
36	Liner					Thermocouple	R-	by Inst. Op.		
37	Type of Heads					Pressure Gauge	R-			
38	Code:					Gauge Glass	R-			
39	Stress Relieve**:		Radiography**:			Level Control	R-			
40	Joint Efficiency:					Safety Valve	R-			
41	Density of Contents:			at	*°F / °C					
42	Weight Empty:			Weight Full:						
43	Is vessel subject to mechanical vibration? *Yes / No					Vent	C-			
44	insulation:	Type: *Frost and Personnel Protection / Cold				Drain	C-			
45	*Yes / No	/ Anticondensation / Heat Conservation				Steam Out	C-			
46	REMARKS:									
47						Manhole	A-			
48										
49										
50										
51										
52										
53						min. Base Elev'n:		k- Skirt Length		
54						Material:				

Issue No.	1	Date	2	Date	3	Date	4	Date
Made/Revised by								
Checked by								
Approved - Process								
Approved by								

TANK SCHEDULE

Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
Job Name:			Description						
Document No.			Made/Revised by						
Project No.			Checked by						
Client			Approved- Process						
Location			Approved						
Plant									

1	Item Number.								
2	Item Name								
3	Number Required								
4	Type	Units							
5	Length x Width	mm	in.						
6	Height	mm	in.						
7	Diameter								
8	Minimum Base Elevation	mm	in.						
9	Total Volume	m^3	ft^3						
10	Working Capacity	m^3	f^{t3}						
11	Internals / Fittings								
12									
13									
14	Operating Pressure	bar g	psig						
15	Operating Temperature	°C	°F						
16	Design Pressure +/ -	bar g	psig						
17	Design Temperature	°C	°F						
18	Vacuum Design								
19	Material: Shell								
20	Liner								
21	Internals								
22									
23	Shell Corrosion Allowance	mm	in.						
24	Sour Service								
25	Stress Relieved for Process Reasons								
26	Insulation								
27									
28									
29									
30	Notes								
31									
32									
33									
34									
35									
36									
37									
38									
39									
40									
41									
42									
43									
44									
45									

	TIE - IN - SCHEDULE								
Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
Job Name:			Description						
Document No.			Made/Revised by						
Project No.			Checked by						
Client			Approved- Process						
Location			Approved						
Plant									

TIE-IN NUMBER	SERVICE	DESCRIPTION	EXISTING LINE NUMBER	NEW LINE NUMBER	EXISTING ELD* NUMBER	NEW ELD* NUMBER	REMARKS	TIE-IN NUMBER

ELD* = ENGINEERING LINE DRAWING

Made by:

Date:

Checked by:

TOWER PROCESS DATA SHEET

Document No.

Job

Item No:(s)

Item Name

Sheet of

Tower Name:

NOTE: * indicates delete as necessary; ** indicates for other than code reason

1				Top	Bottom					
2	Shell Diameters O.D. - I.D.									
3	No. of Trays									
4	Pressure		Operating							
5	*psig		Design							
6	Temperature		Operating							
7	*°F / °C		Design							
8	Material		Shell							
9			Trays							
10			Caps							
11			Liner or Clad							
12	Corrosion -		Shell							
13	Allowance		Heads							
14			Trays							
15	Tray Spacing									
16	Type of Liquid Flow									
17	Type of Trays									
18	Joint Efficiency									
19	Code			Emergency Vac. Design *Yes / No						
20	Stress Relieved **Yes / No			Radiography **Yes / No						
21	Is vessel subject to mechanical vibration *Yes / No									
22	Insulation		Type: *Frost and Personnel Protection / Cold /							
23	*Yes / No		/anticondensation / Heat Conservation /							
24	Min. Base Elevation:				Skirt length:					
25	Weight Empty:				Full:					
26	Nozzles	Mark No.	Size	Number						
27	Feed	C-								
28										
29	Overhead Vpr.	C-								
30	Reflux In	C-								
31										
32	Bottoms	C-								
33	Reboiler Vpr	C-								
34	Reboiler liq	C-								
35										
36										
37	Thermocouple	R-	by inst. Gp.							
38	Level Glass	R-								
39	Press Gauge	R-								
40	Level Control	R-								
41	Safety Valve	R-								
42										
43										
44	Vent	C-								
45	Drain	C-								
46	Steam Out	C-								
47	Manholes	A-								
48	Handholes	A-								
49	Cap Type:			Tray Layout Ref:						
50	NOTES									
51										
52										
53										
54										
55			1	Date	2	Date	3	Date	4	Date
56	Description									
57	Made/Revised by									
58	Checked by									
59	Approved Process									
60	Approved by									

A.K.C. TECHNOLOGY

TRAY LOADING SUMMARY

Job:

Job No.
item No.
Sheet of

Tower Name: Type of Tray

1	Manufacturer				Mfr. Ref:			
2	Pressure at top of Tower	barg	psia					
3	Max. ΔP over Tower	bar	psi		units Used: METRIC/BRITISH (delete one)			
4	No. of Trays			Aboved Feed			Below Feed	
5	Tray location							
6	Tray Number*							
7	Tray Spacing	mm	in.					
8	Tower Internal Diameter I.D.	mm	in.					
9	Vapor to Tray							
10	Temperature	°C	°F					
11	Compressibility							
12	Density	kg/m^3	lb/ft^3					
13	Molecular Weight							
14	Rate	kg/h	lb/h					
15								
16	Liquid from Tray							
17	Temperature	°C	°F					
18	Surface Tension	dynes/cm	lb/ft					
19	Viscosity	cps	lb ft/hr					
20	Density	kg/m^3	lb/ft^3					
21	Rate	kg/h	lb/h					
22	Foaming Tendency **							
23	Number of Passes							
24	Minimum Hole Diameter ***	mm	in.					
25	Minimum DC Residence Time***	secs	s					
26	Maximum Rate as % Design							
27	Minimum Rate as % Design							
28	Design Rate % Flood Rate***							
29								
30	Tray Material							
31	Valve or Cap Material							
32	Corrosion Allowance	mm	in.					
33	Tray Thickness***	mm	in.					
34								
35								
36								
37								
38	NOTES: * Trays are numbered from the bottom of the tower upwardstop of the tower downwards							
39	** Indicate whether 'non', 'moderate', 'high' or 'severe'							
40	*** Data to be supplied by tray manufacturer unless special Process/Client requirement entered here.							
41	REMARKS:							
42								
43								
44								
45								
46								
47								
48								
49								
50								
51								
52								
53								
54								

55			1	Date	2	Date	3	Date
56		Description						
57		Made/Revised by						
58		Checked by						
59		Approve - Process						
60		Approved by						

TRIP SCHEDULE

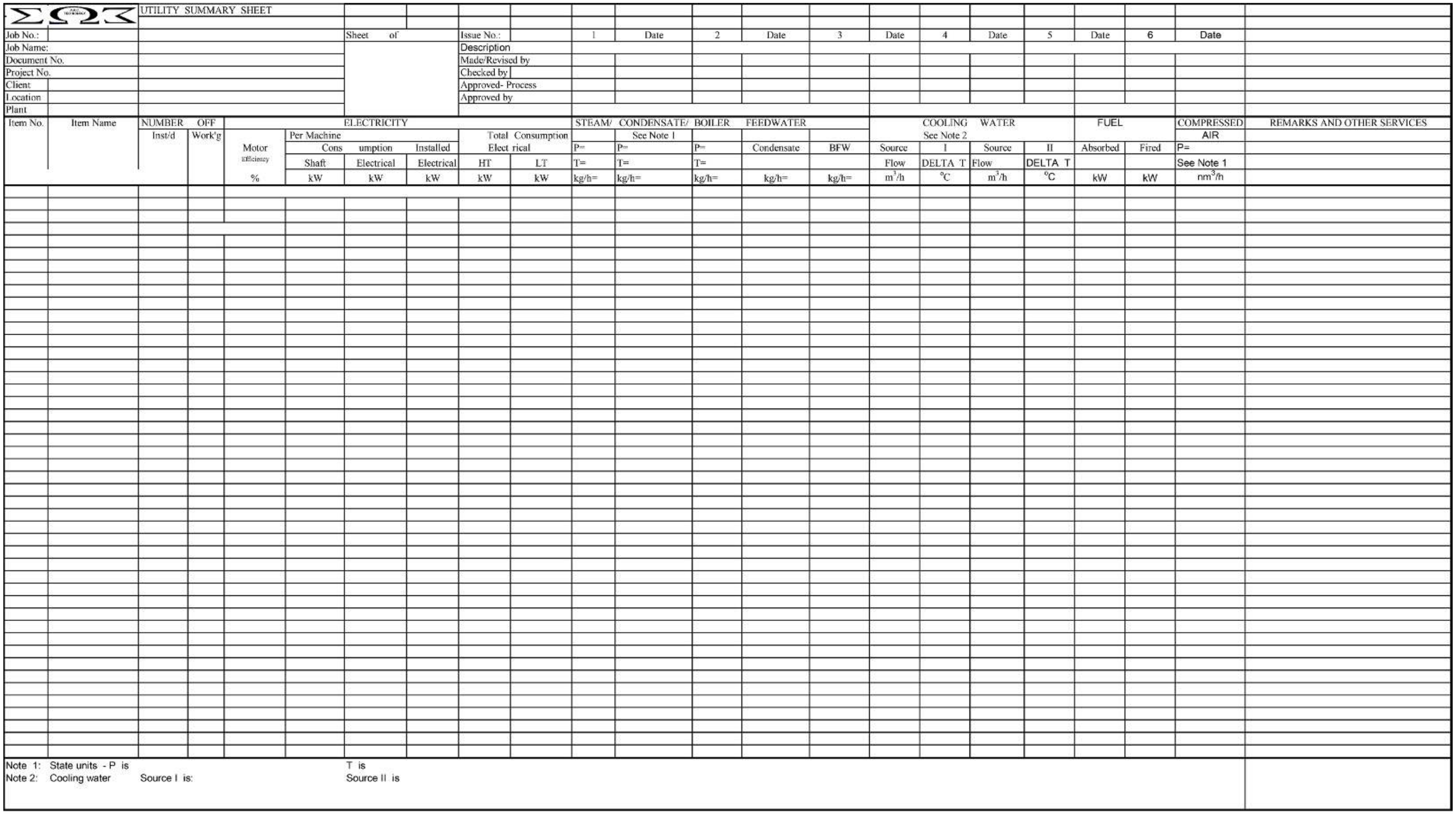
Job No.:
Job Name:
Document No.
Project No.
Client
Location
Plant

Sheet of

Issue No.	1	Date	2	Date
Decription				
Made/Revised by				
Checked by				
Approved Process				
Approved				

1	2	3	4	5	6	7	8	9	10	11	12
INITIATING VARIABLE	BASIC	PRE-	ALARM	CUT	OUT	FIRST FAULT	INHIBIT	TRIP LINKAGE	ITEM	PRIMARY	REMARKS
	INST. No.	INST. No.	SETTING	INST. No.	SETTING	INDICATION	SWITCH		ACTUATED	PROCESS	
										EFFECT	

Trip Schedule.



UTILITY SUMMARY SHEET

Job No.:	Sheet of	Issue No.:	1	Date	2	Date	3	Date	4	Date	5	Date	6	Date
Job Name:		Description												
Document No.		Made/Revised by												
Project No.		Checked by												
Client		Approved- Process												
Location		Approved by												
Plant														

Item No.	Item Name	NUMBER	OFF	ELECTRICITY						STEAM/	CONDENSATE/	BOILER	FEEDWATER		COOLING	WATER			FUEL		COMPRESSED	REMARKS AND OTHER SERVICES
		Inst/d	Work'g		Per Machine			Total Consumption			See Note 1				See Note 2						AIR	
				Motor	Cons	umption	Installed	Elect rical		P=	P=	P=	Condensate	BFW	Source	I	Source	II	Absorbed	Fired	P=	
				Efficiency	Shaft	Electrical	Electrical	HT	LT	T=	T=	T=			Flow	DELTA T	Flow	DELTA T			See Note 1	
				%	kW	kW	kW	kW	kW	kg/h=	kg/h=	kg/h=	kg/h=	kg/h=	m³/h	°C	m³/h	°C	kW	kW	nm³/h	

Note 1: State units - P is T is

Note 2: Cooling water Source I is: Source II is

Utility summary sheet.

VESSEL AND TANK SCHEDULE									
Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
Job Name:			Description						
Document No.			Made/Revised by						
Project No.			Checked by						
Client			Approved- Process						
Location			Approved						
Plant									

ITEM No.	No. UNITS	SERVICE DESCRIPTION	VERTICAL OR HORIZONTAL	DIAMETER	LENGTH	THICKNESS		LINING	INSULATION CODE	TEST PRESSURE, PSIG	WEIGHT	REFERENCE DRAWING	PURCHASE ORDER TO	P.O. Number
						HEAD	SHELL							

Vessel and tank schedule.

VESSEL AND TANK SUMMARY
Driver

Job No.:
Job Name:
Document No.
Project No.
Client
Location
Plant

Sheet of

Issue No.:	1	Date	2	Date	3	Date	4	Date	5	Date
Description										
Made/Revised by										
Checked by										
Approved- Process										
Approved										

				CAPACITY		THICKNESS, INCHES			CORROSION	MATERIAL		PRESSURE, PSIG			TEMP, °F					
ITEM No.	EQUIPMENT	TYPE	O.D.	HEIGHT or LENGTH.	GALLONS	SHELL	HEAD	SADDLE SKIRT SUPPORT	ALLOWANCE INCHES	SHELL & HEADS	SADDLE SKIRT SUPPORT	OPER'N	DESIGN	TEST	OPER'N	DESIGN	CODE	STAMP	REFERENCE DRAWING No.	NOTES

Vessel and tank summary

VESSEL SCHEDULE

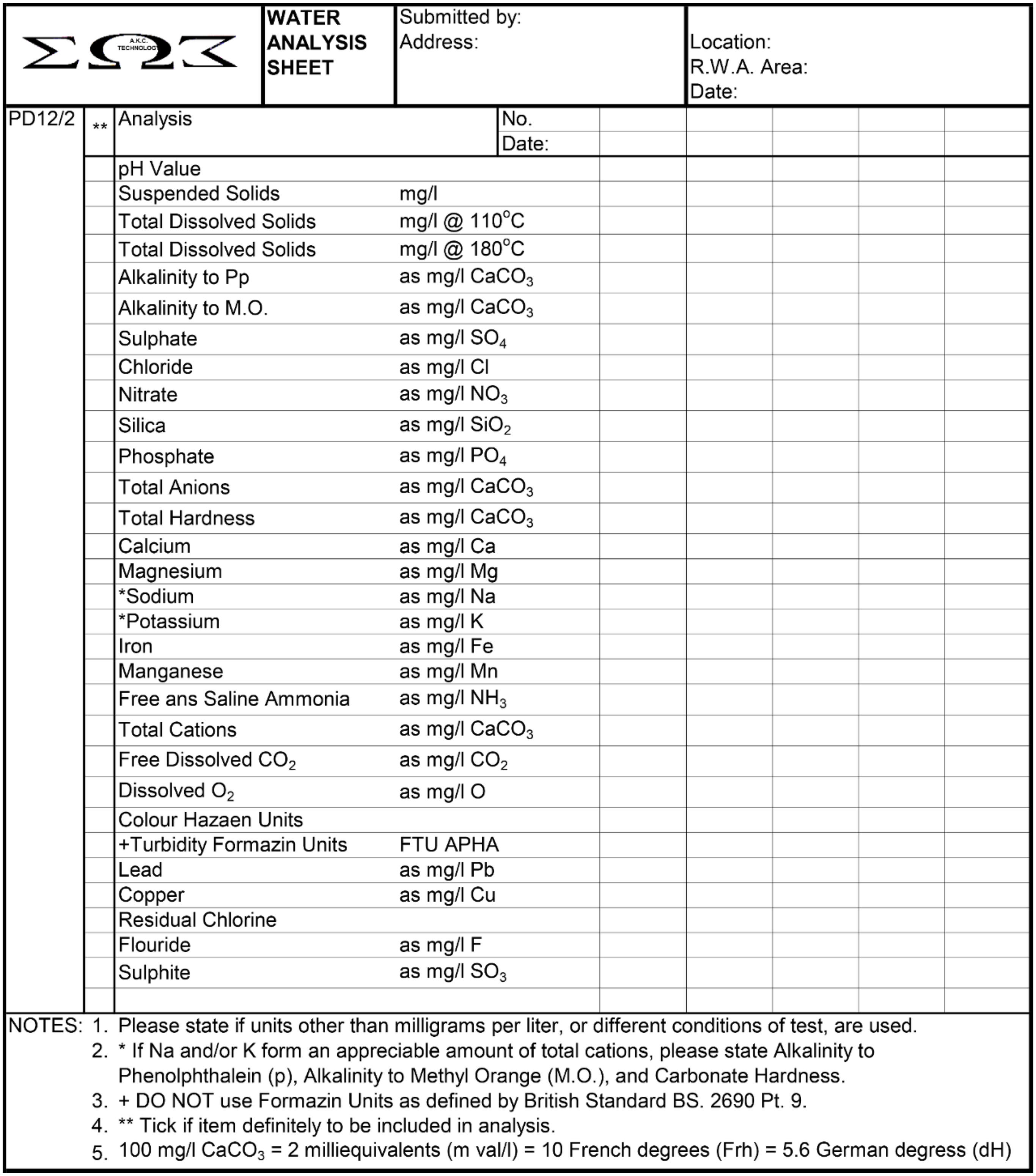
Job No.:		Sheet of	Issue No.:	1	Date	2	Date	3	Date
Job Name:			Description						
Document No.			Made/Revised by						
Project No.			Checked by						
Client			Approved- Process						
Location			Approved						
Plant									

1	Item Number.									
2	Item Name									
3	Number Required									
4		Units								
5	Shell Diameter (O.D/I.D.)	mm	in.							
6	Shell Length	mm	in.							
7	Center Line (H/V)									
8	Boot Size	mm	in.							
9	Base Elevation	mm	in.							
10	Internals									
11										
12										
13										
14	Operating Pressure	bar g	psig							
15	Operating Temperature	°C	°F							
16	Design Pressure	bar g	psig							
17	Design Temperature	°C	°F							
18	Vacuum Design									
19	Material: Shell									
20	Liner									
21	Internals									
22										
23	Shell Corrosion Allowance	mm	in.							
24	Sour Service									
25	Stress Relieved for Process Reasons									
26	Insulation									
27										
28										
29										
30	Notes									
31										
32										
33										
34										
35										
36										
37										
38										
39										
40										
41										
42										
43										
44										
45										

A.K.C. TECHNOLOGY	**WATER ANALYSIS SHEET**	Submitted by: Address:	Location: Date:

**	Analysis	No.					
		Date:					
	pH Value						
	Suspended Solids	mg/l					
	Total Dissolved Solids	mg/l @ 110°C					
	Total Dissolved Solids	mg/l @ 180°C					
	Alkalinity to Pp	as mg/l $CaCO_3$					
	Alkalinity to M.O.	as mg/l $CaCO_3$					
	Sulphate	as mg/l SO_4					
	Chloride	as mg/l Cl					
	Nitrate	as mg/l NO_3					
	Silica	as mg/l SiO_2					
	Phosphate	as mg/l PO_4					
	Total Anions	as mg/l $CaCO_3$					
	Total Hardness	as mg/l $CaCO_3$					
	Calcium	as mg/l Ca					
	Magnesium	as mg/l Mg					
	*Sodium	as mg/l Na					
	*Potassium	as mg/l K					
	Iron	as mg/l Fe					
	Manganese	as mg/l Mn					
	Free and Saline Ammonia	as mg/l NH_3					
	Total Cations	as mg/l $CaCO_3$					
	Free Dissolved CO_2	as mg/l CO_2					
	Dissolved O_2	as mg/l O					
	Colour Hazaen Units						
	+Turbidity Formazin Units	FTU APHA					
	Lead	as mg/l Pb					
	Copper	as mg/l Cu					
	Residual Chlorine						
	Flouride	as mg/l F					
	Sulfite	as mg/l SO_3					

NOTES:
1. Please state if units other than milligrams per liter, or different conditions of test, are used.
2. * If Na and/or K form an appreciable amount of total cations, please state Alkalinity to Phenolphthalein (p), Alkalinity to Methyl Orange (M.O.), and Carbonate Hardness.
3. + DO NOT use Formazin Units as defined by British Standard BS. 2690 Pt. 9.
4. ** Tick if item definitely to be included in analysis.
5. 100 mg/l $CaCO_3$ = 2 milliequivalents/ (m val/l)



WATER ANALYSIS SHEET

Submitted by:
Address:

Location:
R.W.A. Area:
Date:

PD12/2	**	Analysis	No.					
			Date:					
		pH Value						
		Suspended Solids	mg/l					
		Total Dissolved Solids	mg/l @ 110°C					
		Total Dissolved Solids	mg/l @ 180°C					
		Alkalinity to Pp	as mg/l $CaCO_3$					
		Alkalinity to M.O.	as mg/l $CaCO_3$					
		Sulphate	as mg/l SO_4					
		Chloride	as mg/l Cl					
		Nitrate	as mg/l NO_3					
		Silica	as mg/l SiO_2					
		Phosphate	as mg/l PO_4					
		Total Anions	as mg/l $CaCO_3$					
		Total Hardness	as mg/l $CaCO_3$					
		Calcium	as mg/l Ca					
		Magnesium	as mg/l Mg					
		*Sodium	as mg/l Na					
		*Potassium	as mg/l K					
		Iron	as mg/l Fe					
		Manganese	as mg/l Mn					
		Free ans Saline Ammonia	as mg/l NH_3					
		Total Cations	as mg/l $CaCO_3$					
		Free Dissolved CO_2	as mg/l CO_2					
		Dissolved O_2	as mg/l O					
		Colour Hazaen Units						
		+Turbidity Formazin Units	FTU APHA					
		Lead	as mg/l Pb					
		Copper	as mg/l Cu					
		Residual Chlorine						
		Flouride	as mg/l F					
		Sulphite	as mg/l SO_3					

NOTES:
1. Please state if units other than milligrams per liter, or different conditions of test, are used.
2. * If Na and/or K form an appreciable amount of total cations, please state Alkalinity to Phenolphthalein (p), Alkalinity to Methyl Orange (M.O.), and Carbonate Hardness.
3. + DO NOT use Formazin Units as defined by British Standard BS. 2690 Pt. 9.
4. ** Tick if item definitely to be included in analysis.
5. 100 mg/l $CaCO_3$ = 2 milliequivalents (m val/l) = 10 French degrees (Frh) = 5.6 German degress (dH)

Appendix F

Equilibrium K-Values

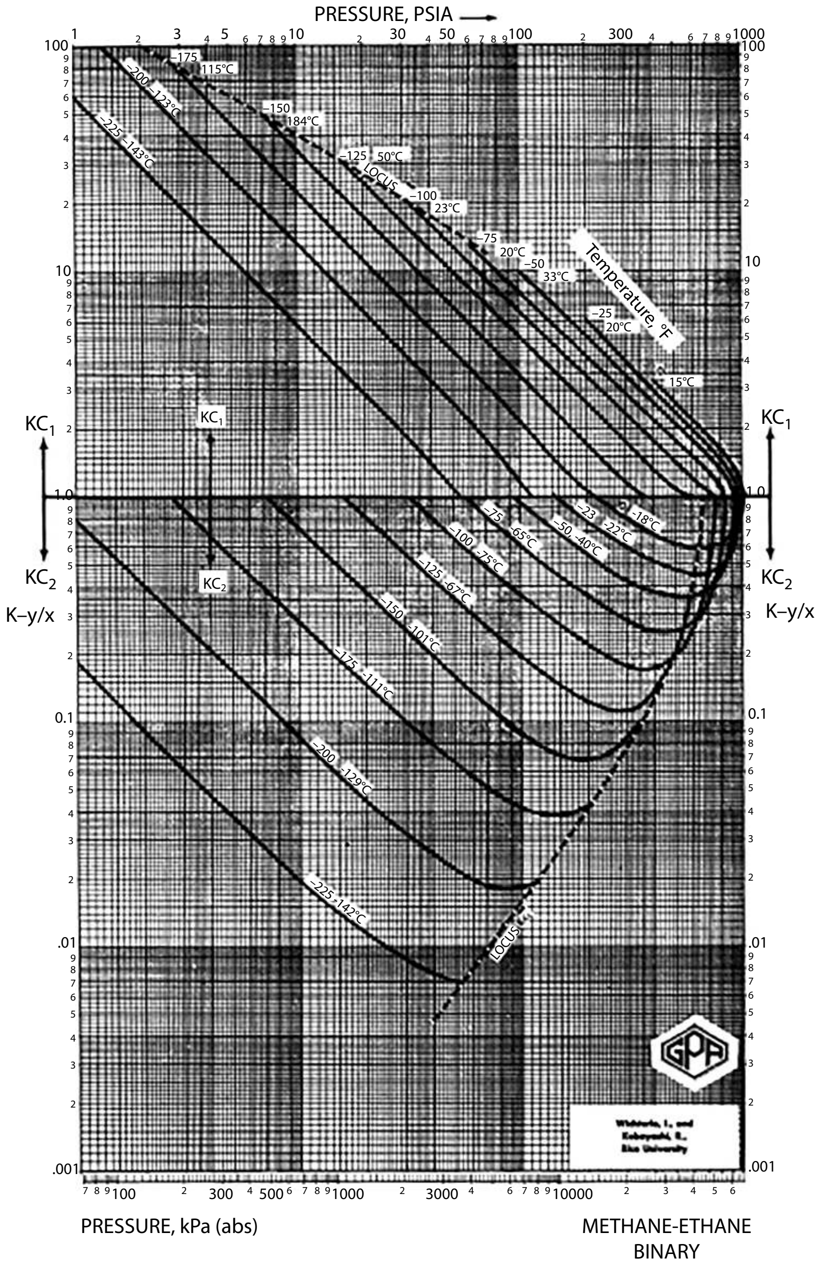


Figure F.1 Pressure vs. K for Methane-Ethane binary. Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

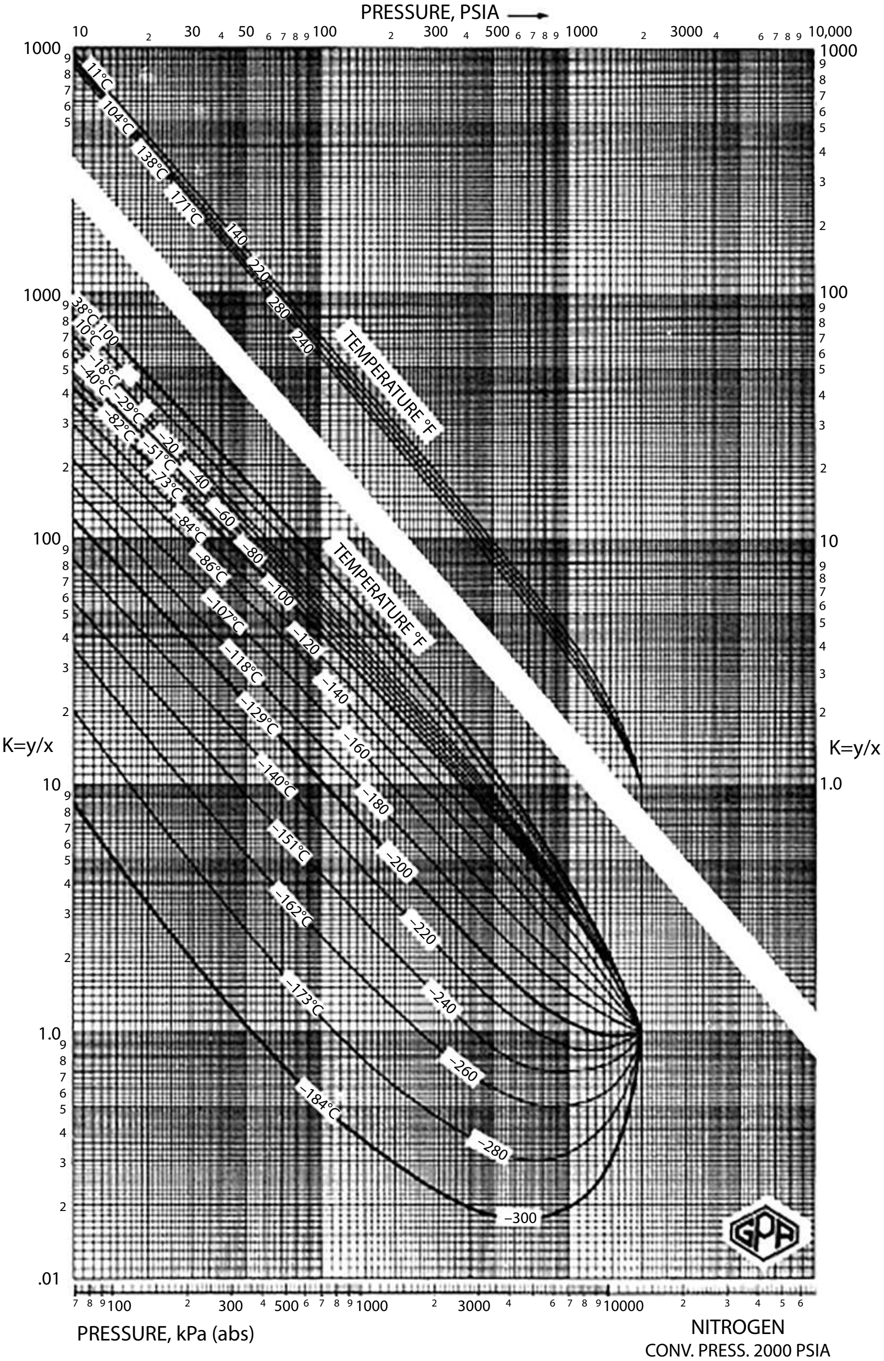


Figure F.2 Pressure vs. K for Nitrogen at convergence pressure 2000 psia (13,800 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

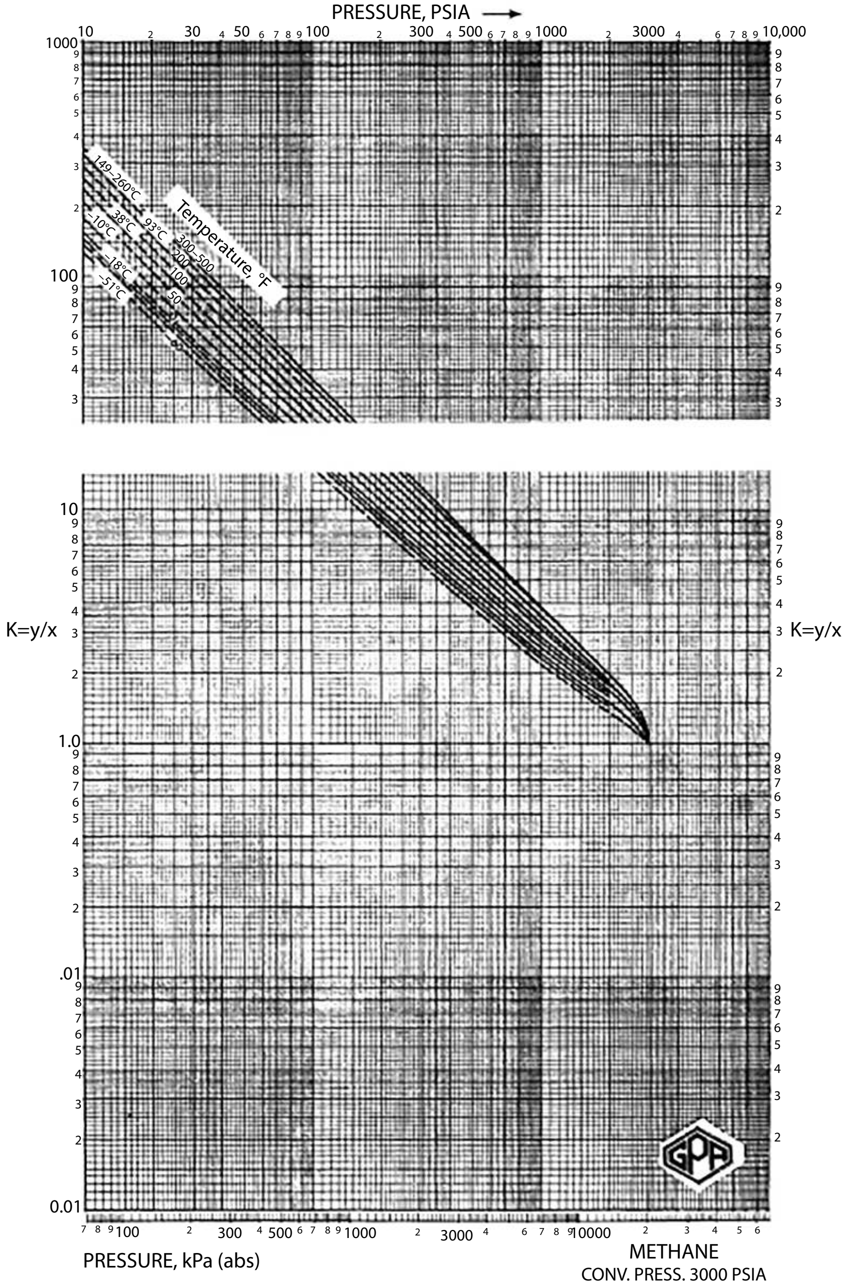


Figure F.3 Pressure vs. K for Methane (CH_4) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

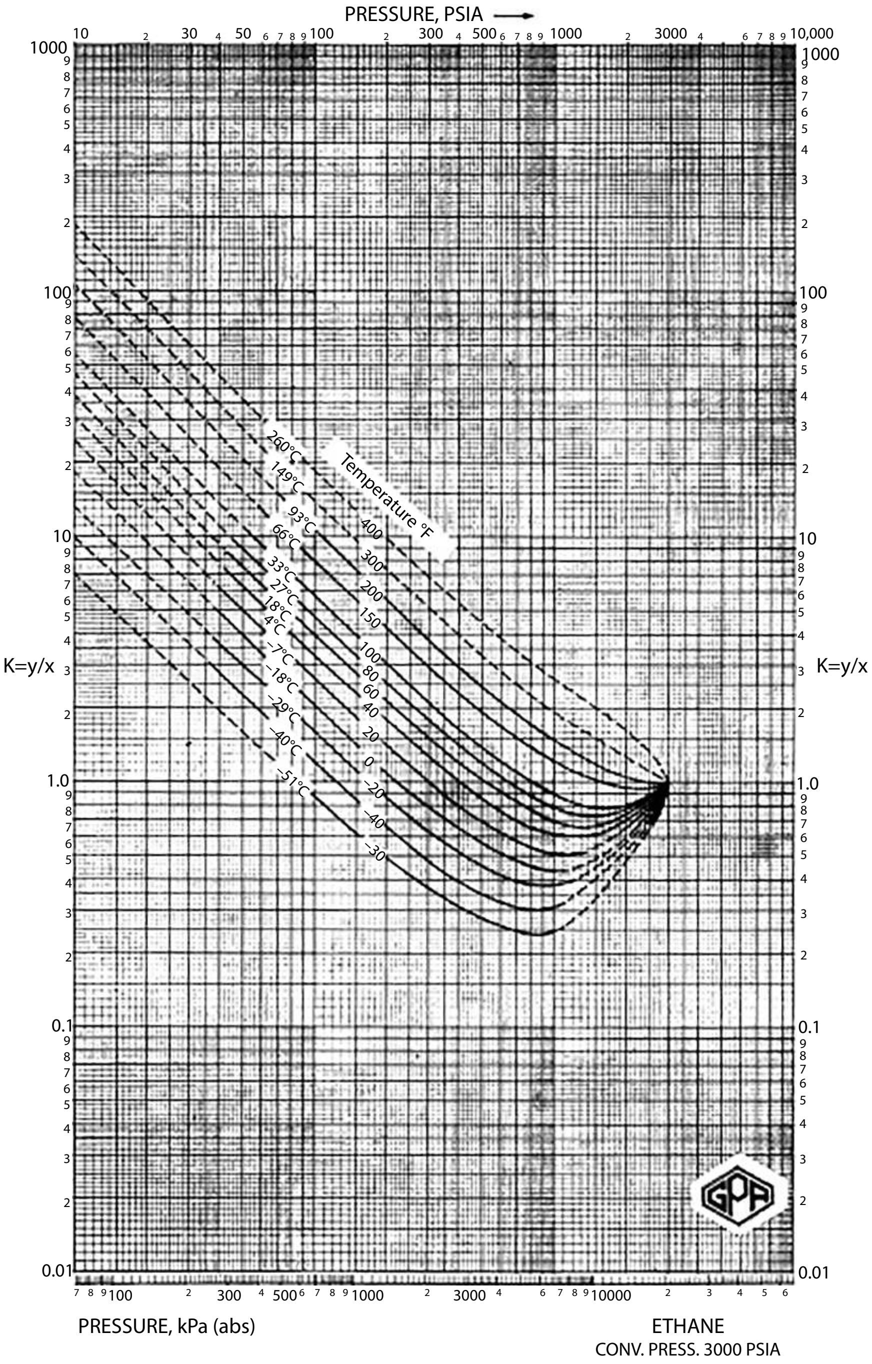


Figure F.4 Pressure vs. K for Ethane (C_2H_6) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

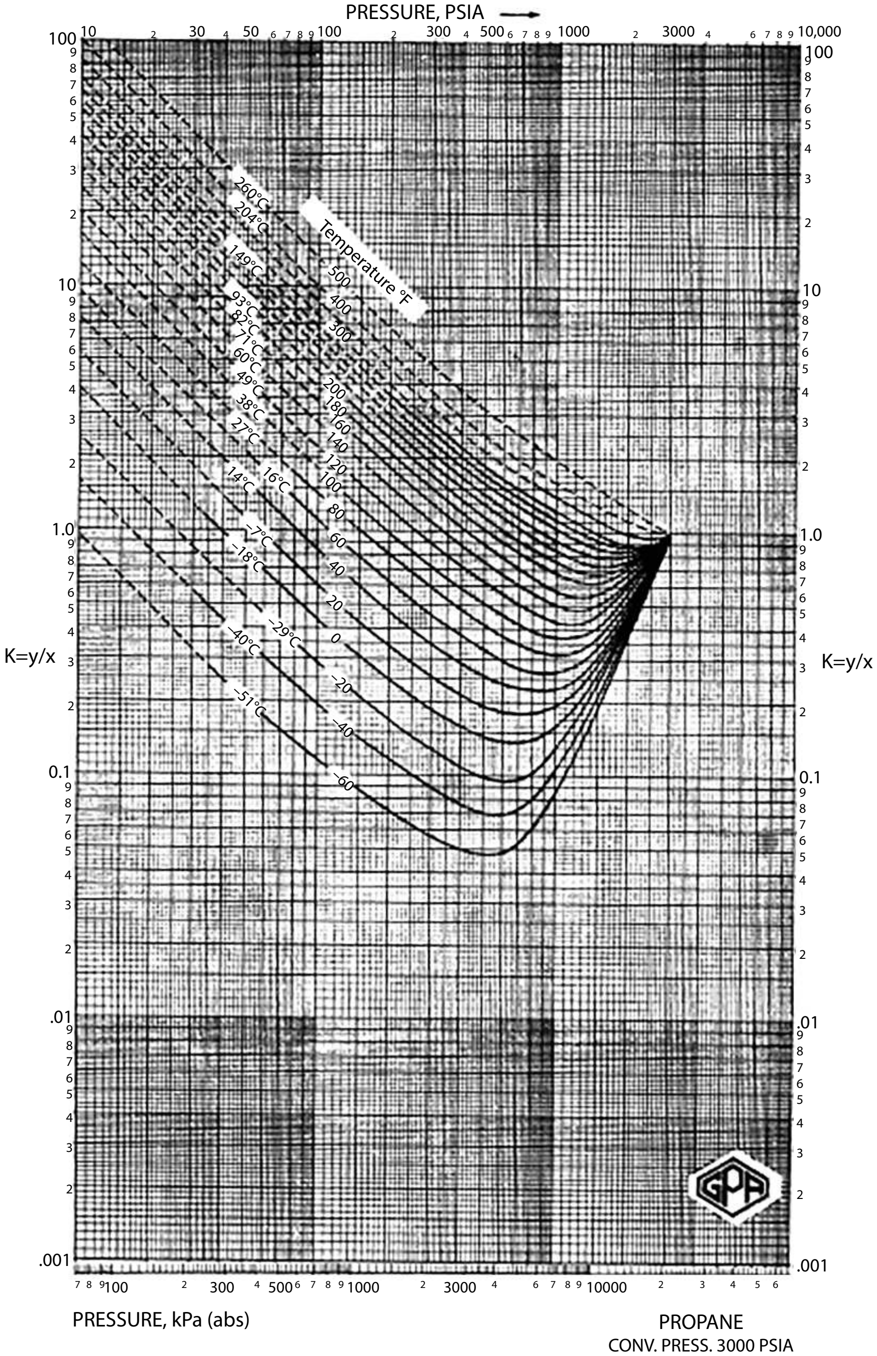


Figure F.5 Pressure vs. K for Propane (C_3H_8) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

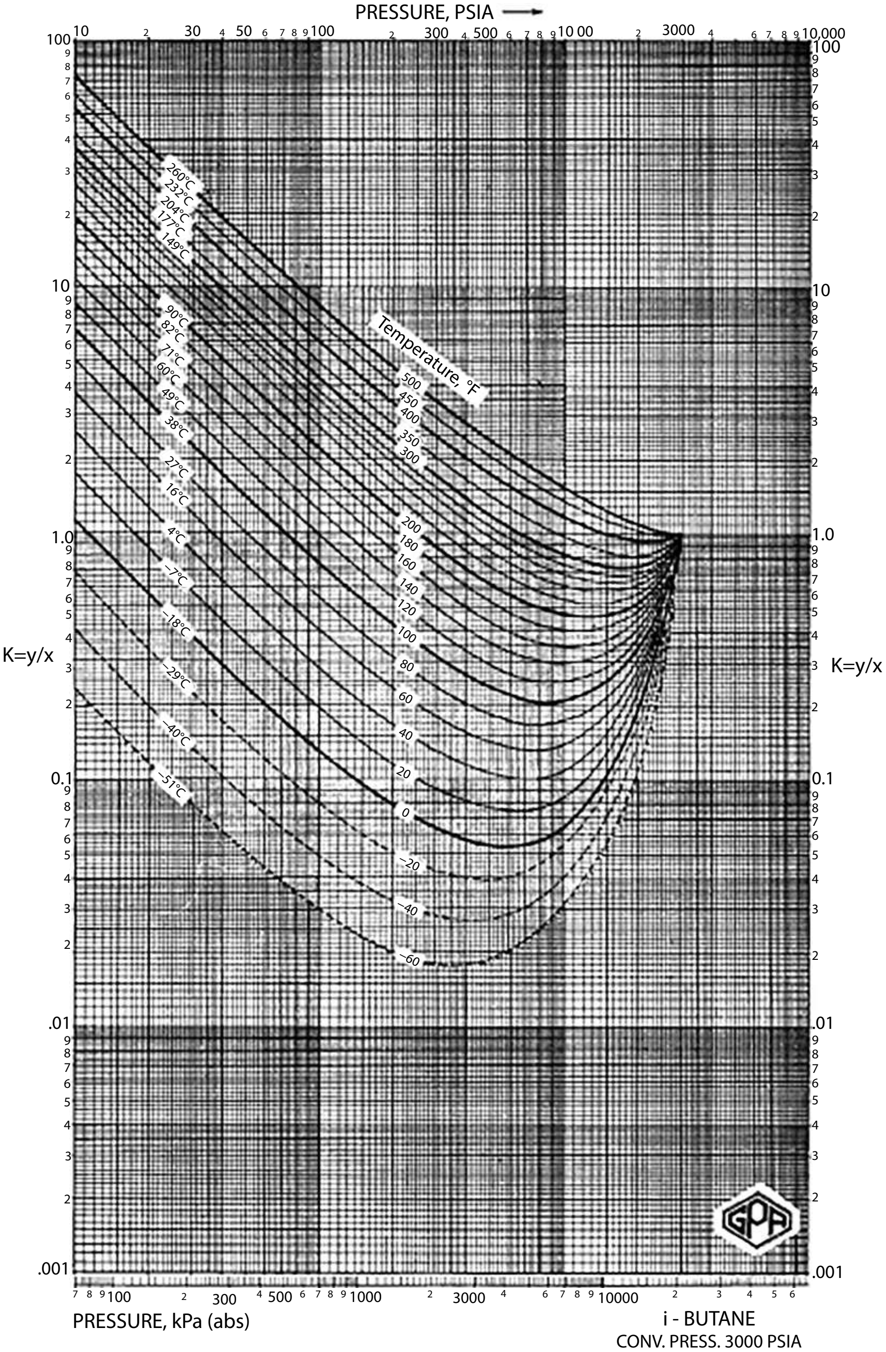


Figure F.6 Pressure vs. K for i-Butane (iC_4H_{10}) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

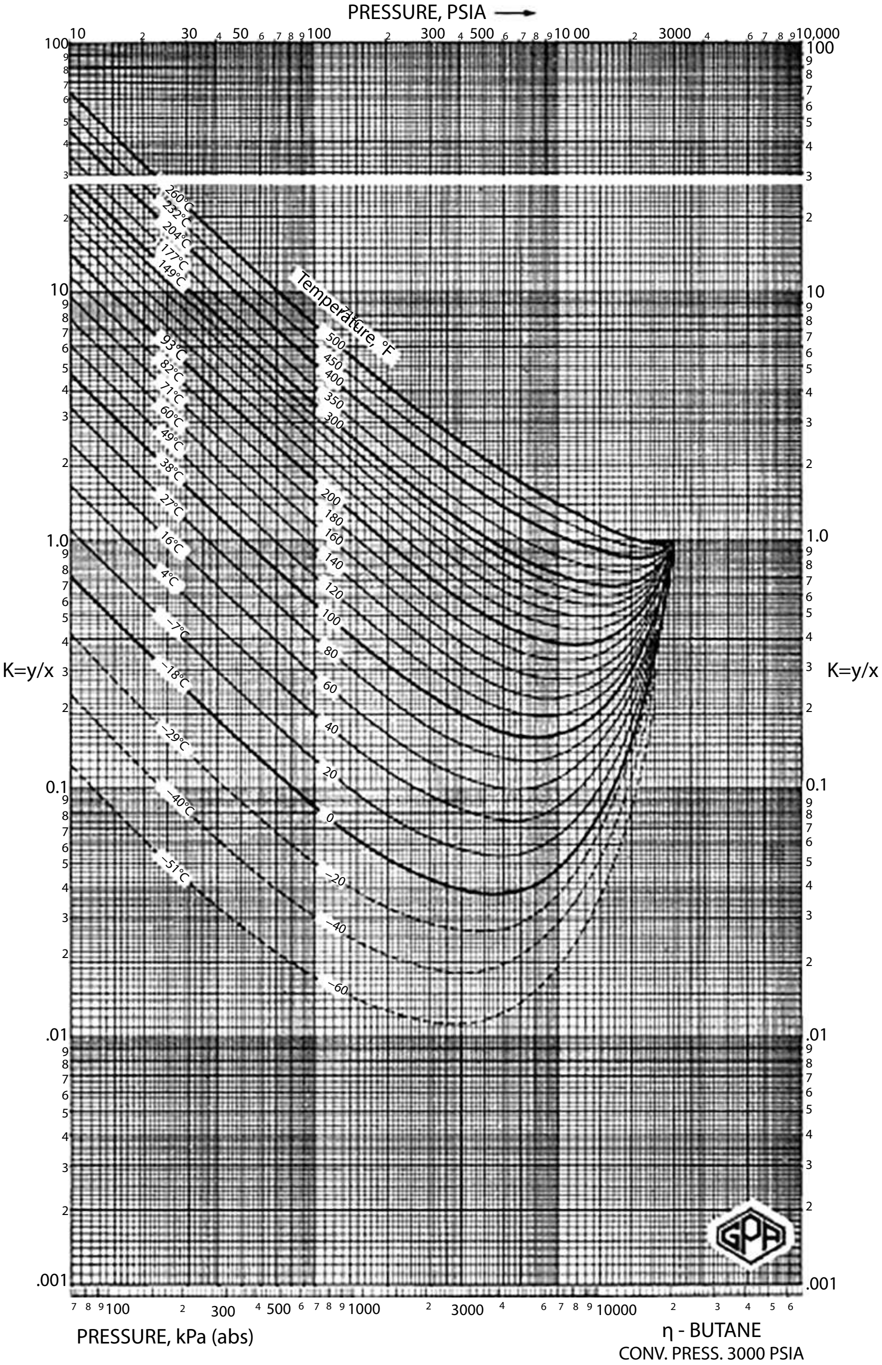


Figure F.7 Pressure vs. K for n-Butane (nC_4H_{10}) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

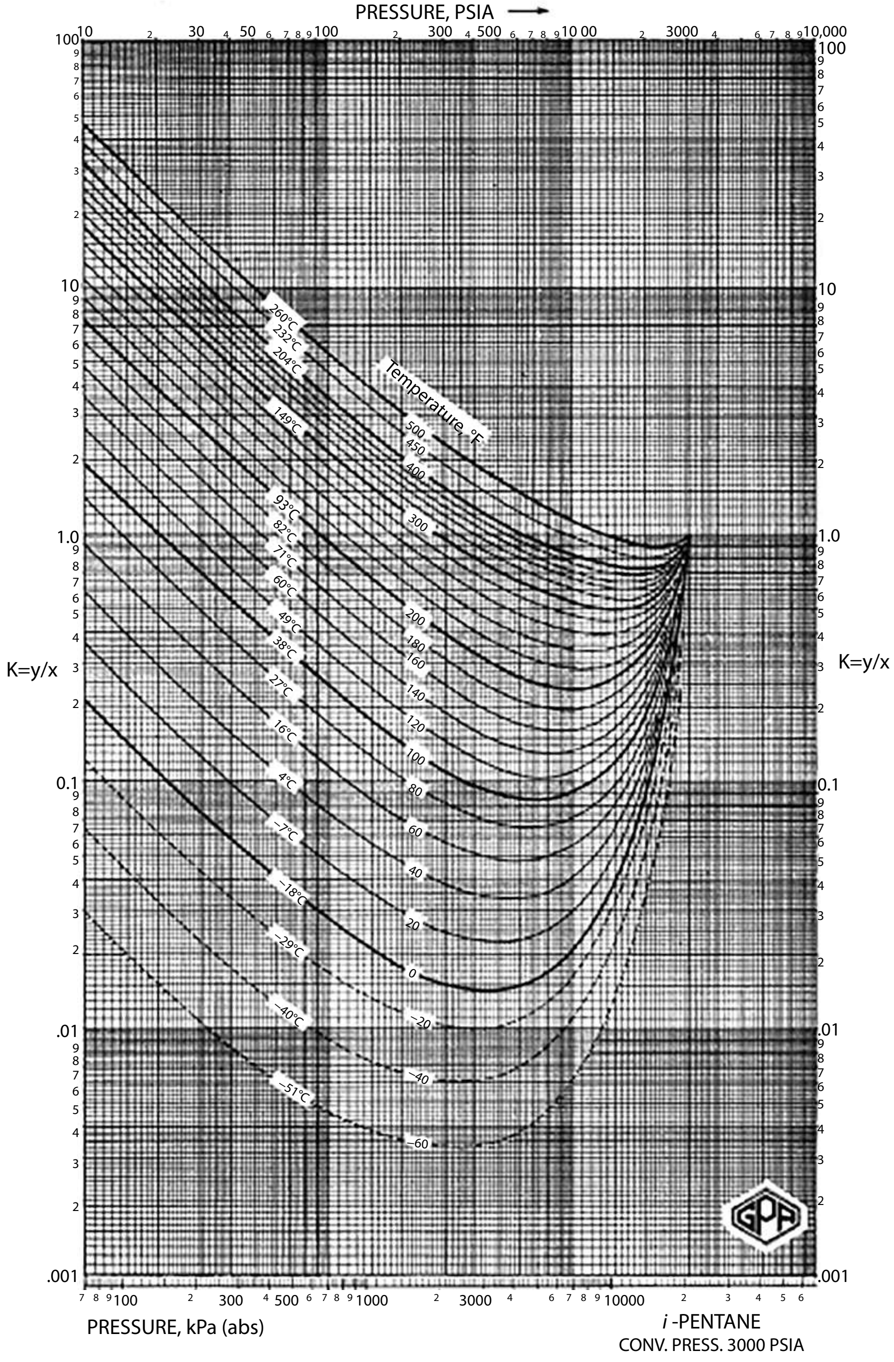


Figure F.8 Pressure vs. K for i-Pentane (iC_5H_{12}) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

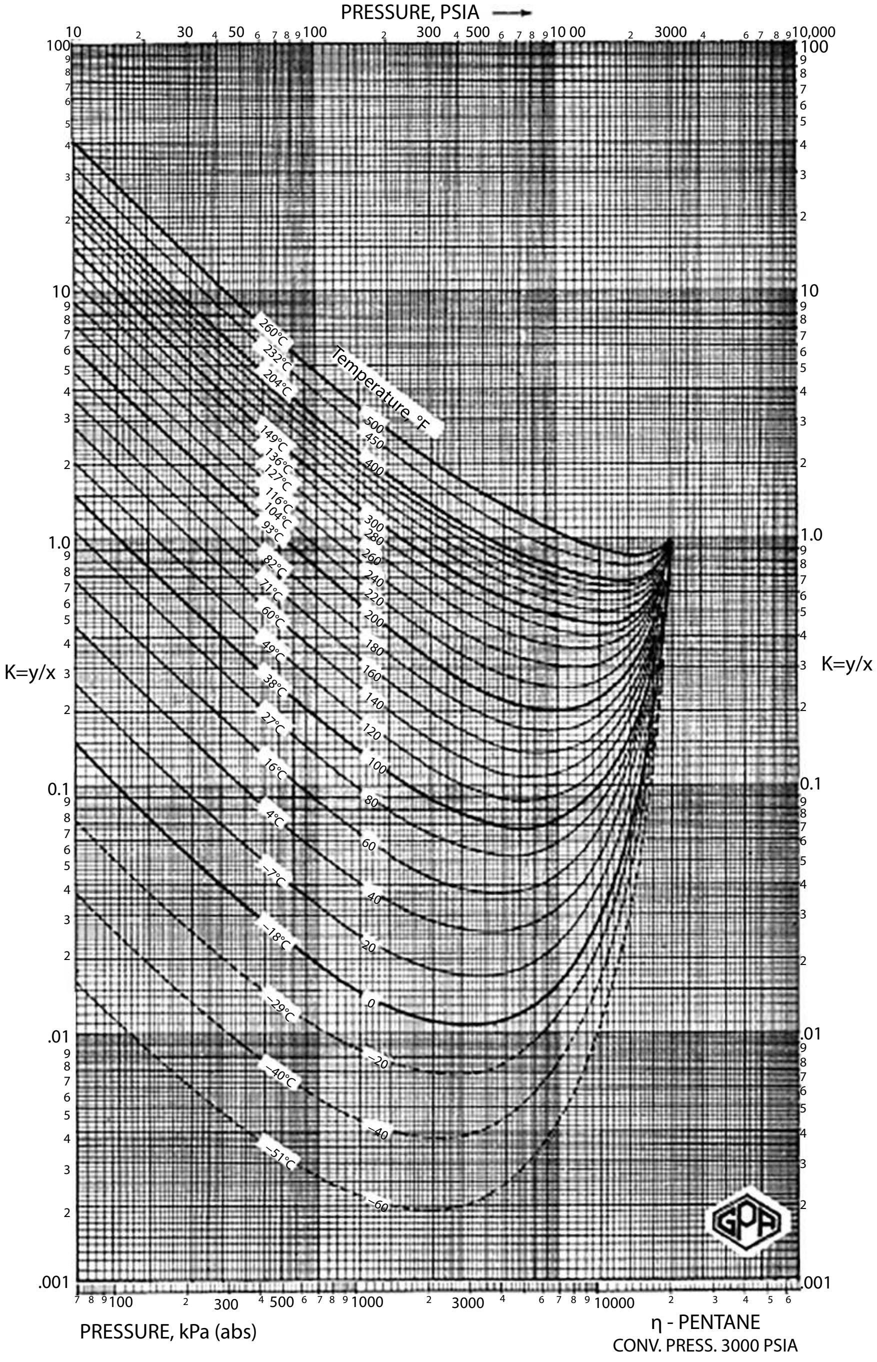


Figure F.9 Pressure vs. K for n-Pentane (nC_5H_{12}) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

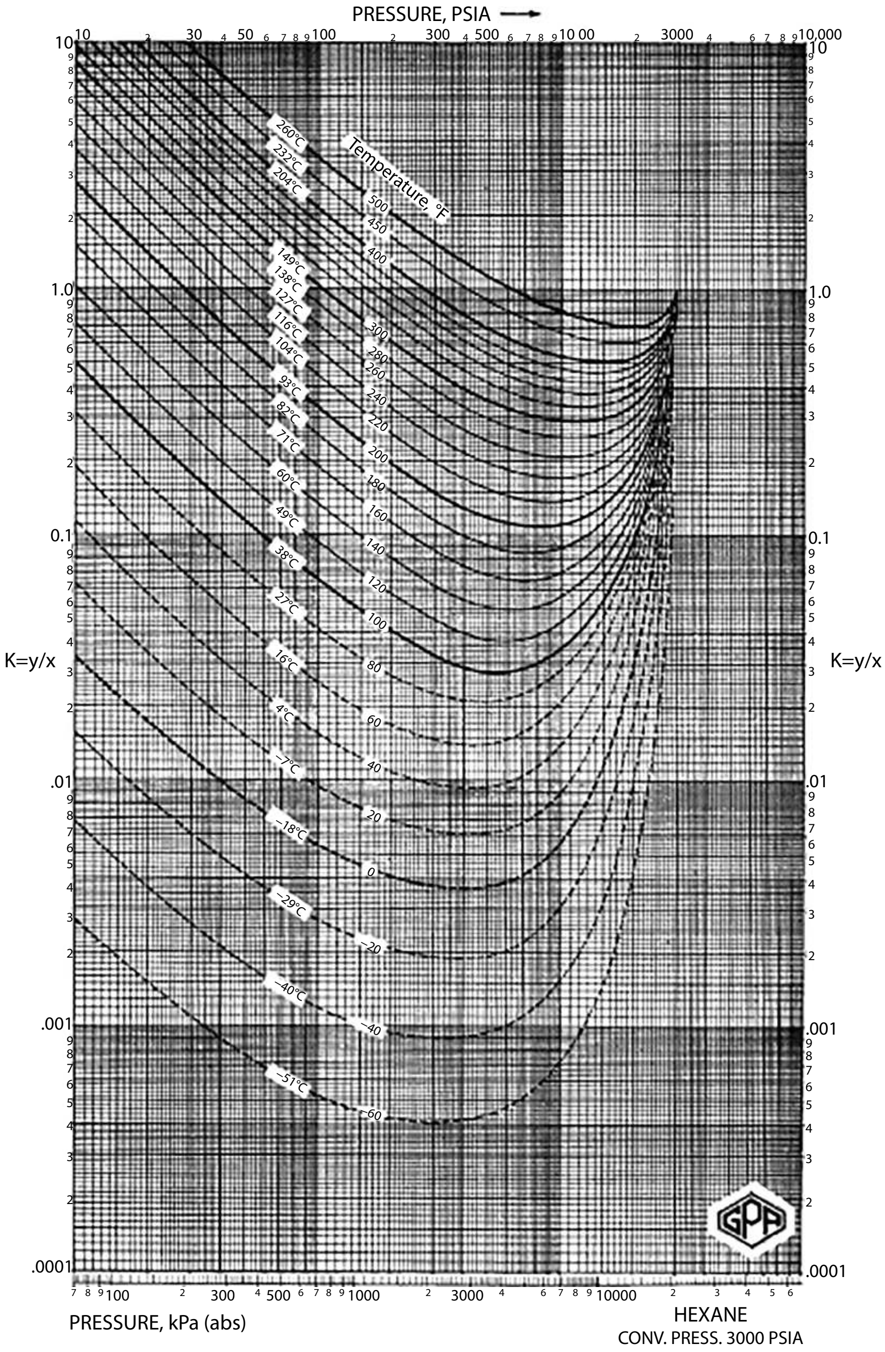


Figure F.10 Pressure vs. K for Hexane (C_6H_{14}) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

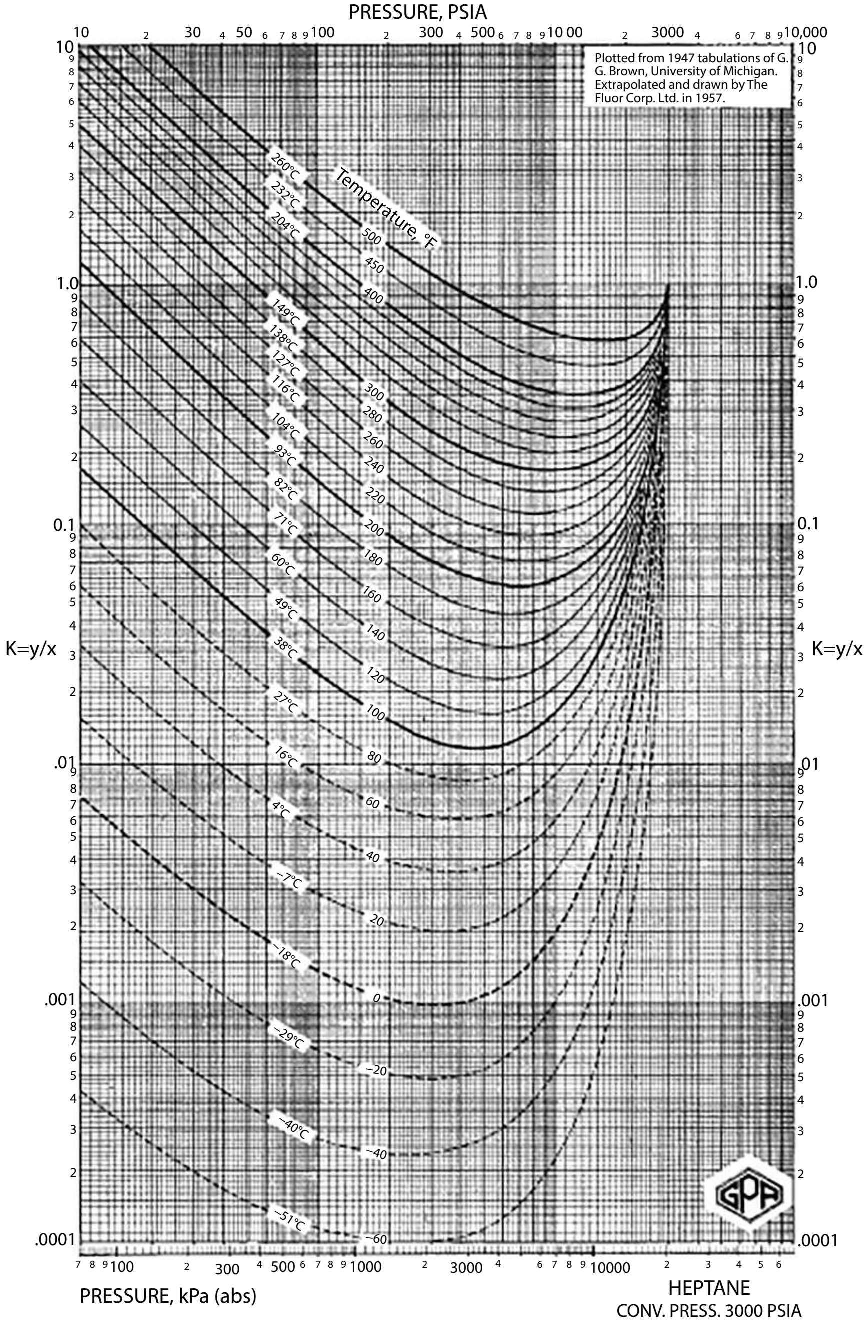

Plotted from 1947 tabulations of G. G. Brown, University of Michigan. Extrapolated and drawn by The Fluor Corp. Ltd. in 1957.

Figure F.11 Pressure vs. K for Heptane (C_7H_{16}) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

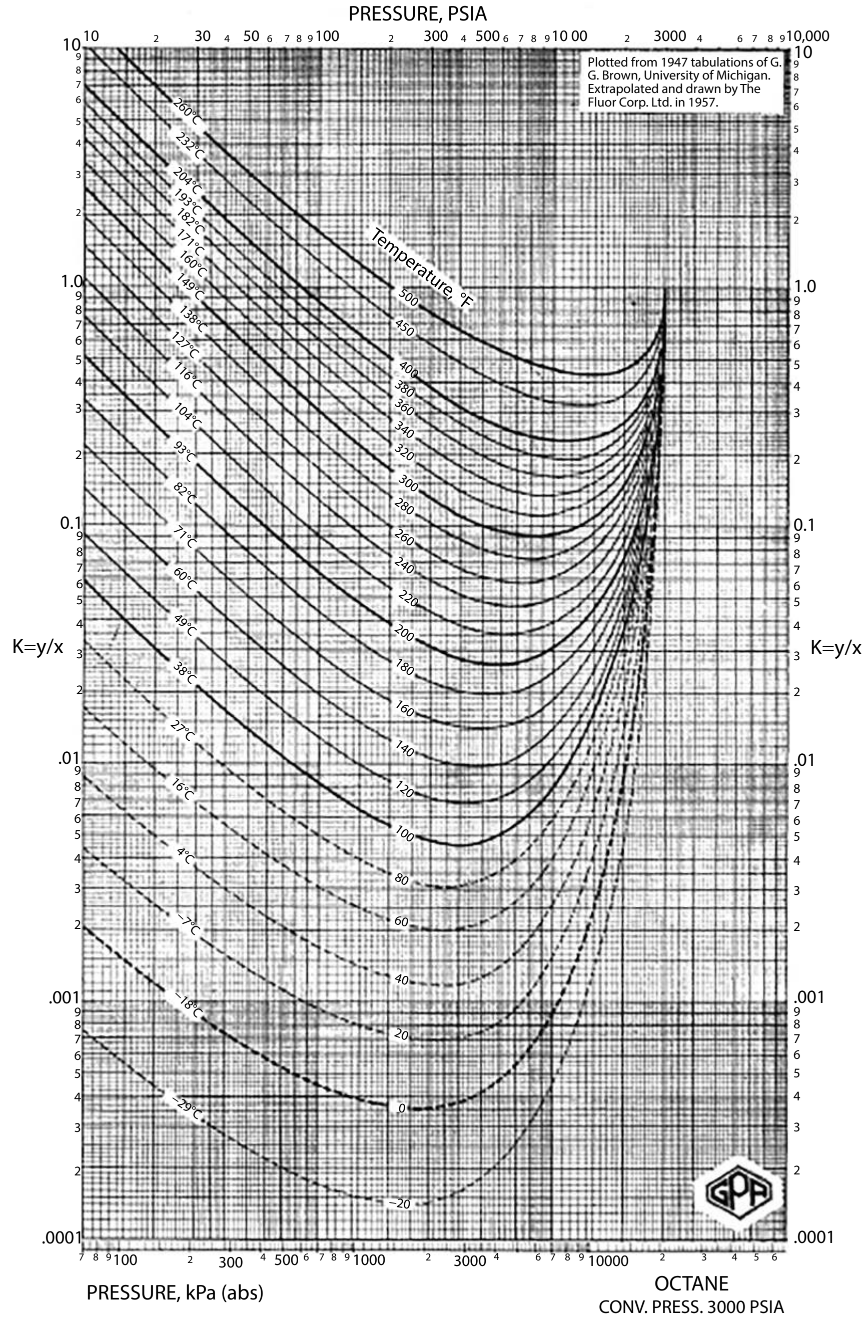
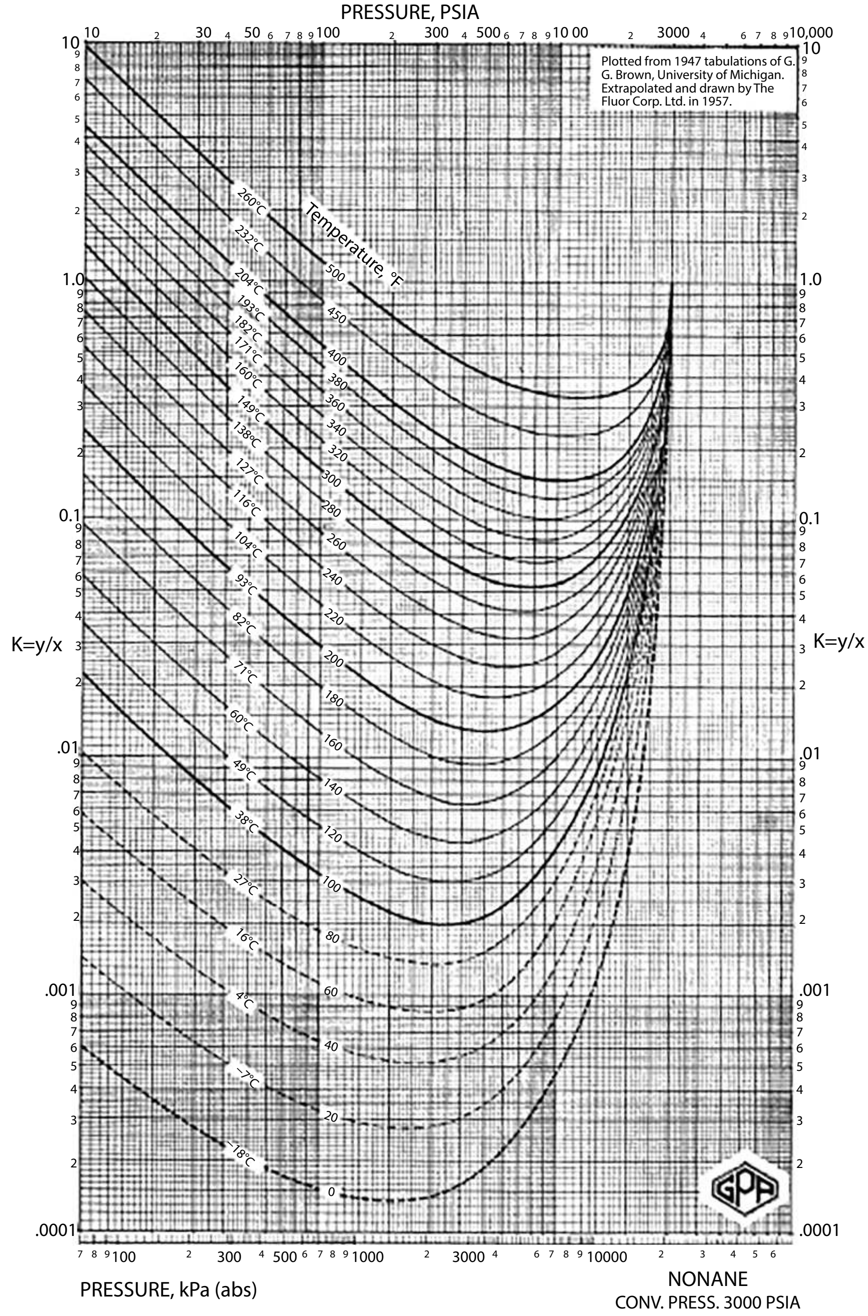

Plotted from 1947 tabulations of G. G. Brown, University of Michigan. Extrapolated and drawn by The Fluor Corp. Ltd. in 1957.

Figure F.12 Pressure vs. K for Octane (C_8H_{18}) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.



Plotted from 1947 tabulations of G. G. Brown, University of Michigan. Extrapolated and drawn by The Fluor Corp. Ltd. in 1957.

Figure F.13 Pressure vs. K for Nonane (C_9H_{20}) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

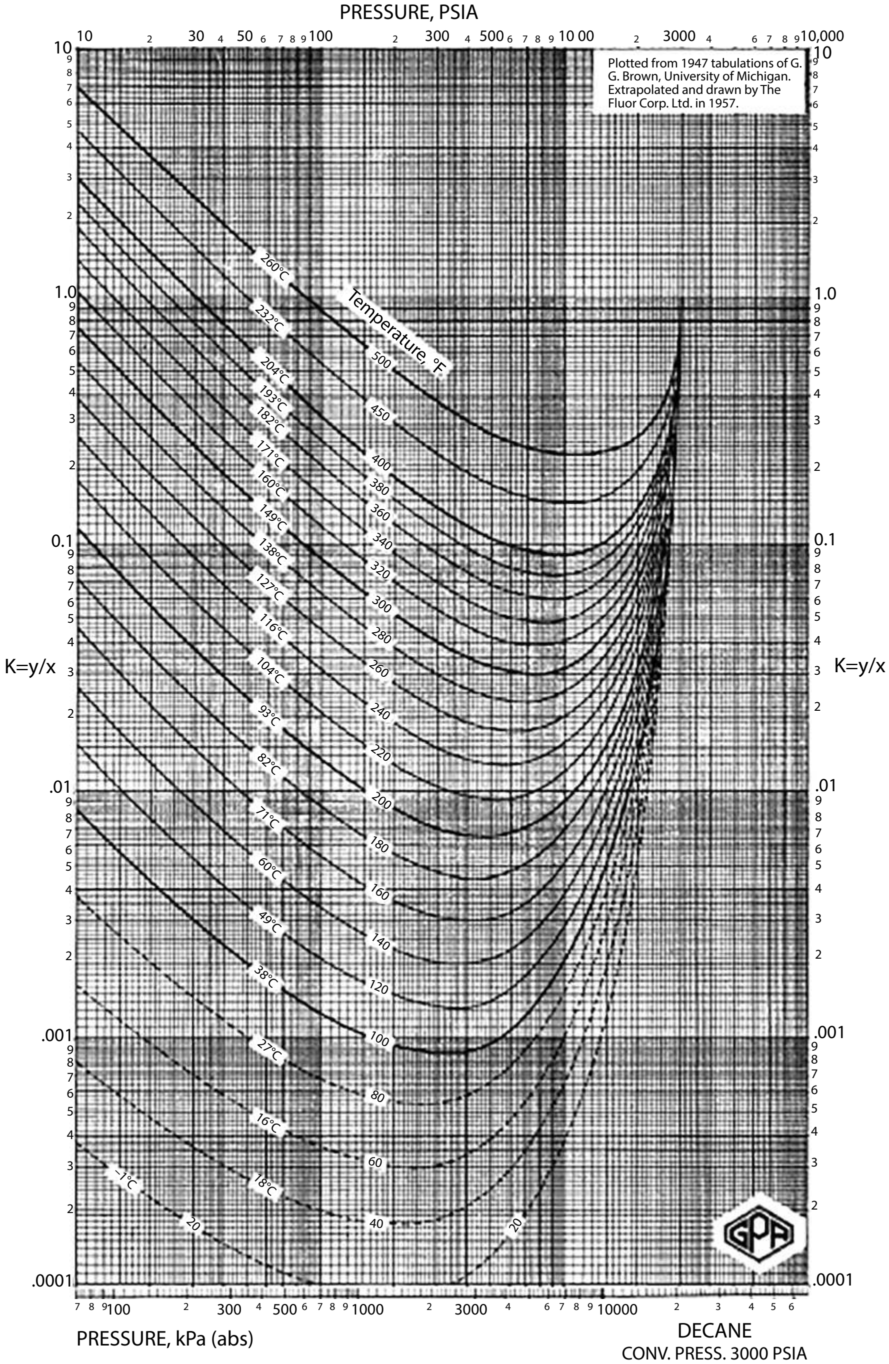


Figure F.14 Pressure vs. K for Decane ($C_{10}H_{22}$) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

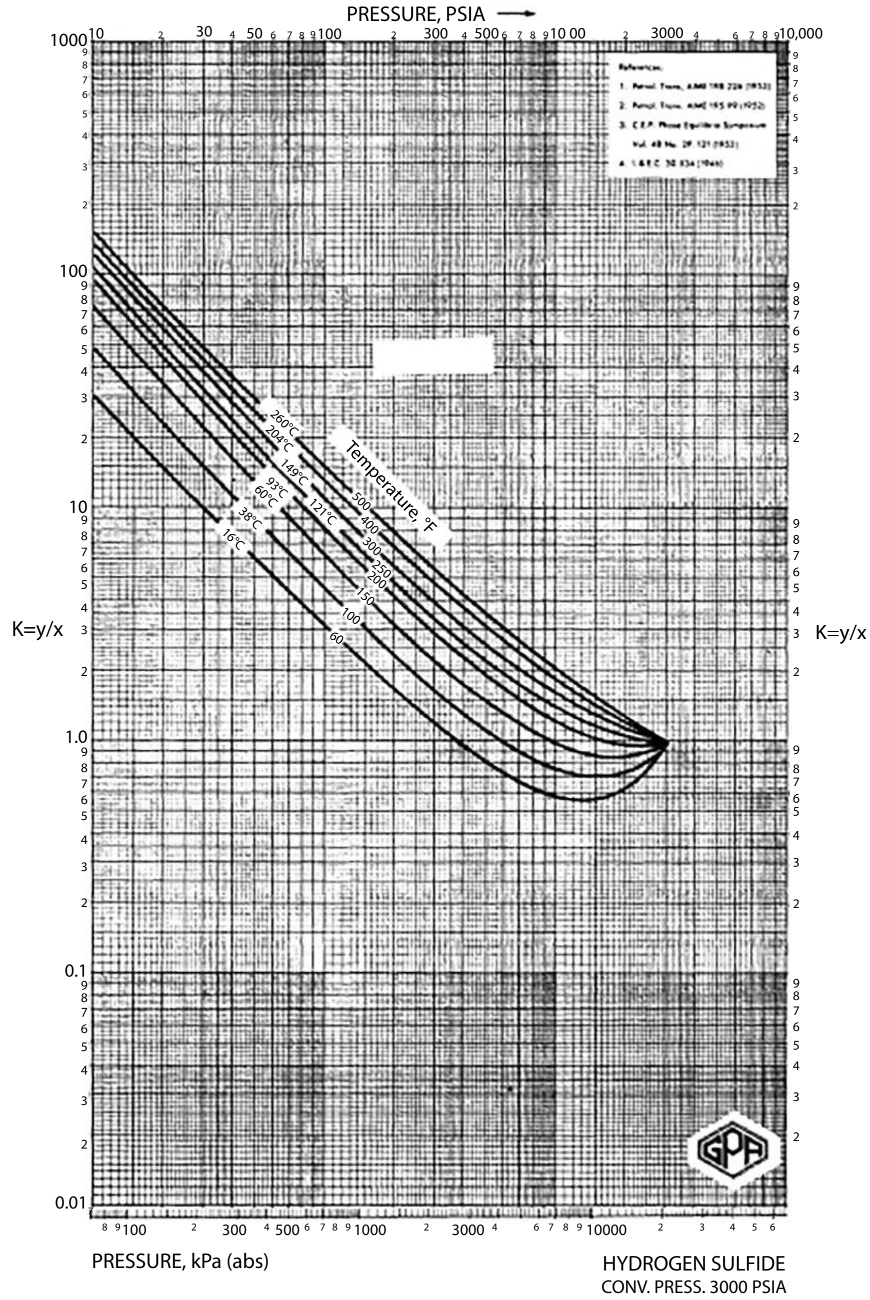


Figure F.15 Pressure vs. K for Hydrogen sulfide (H_2S) at convergence pressure 3000 psia (20,700 kPa). Used by permission, Gas Processors Suppliers Association Data Book, 12th Ed., V. 1 and 2 (2004), Tulsa, Okla.

About the Author

A. Kayode Coker PhD, is Engineering Consultant for AKC Technology, an Honorary Research Fellow at the University of Wolverhampton, U.K., a former Engineering Coordinator at Saudi Aramco Shell Refinery Company (SASREF) and Chairman of the department of Chemical Engineering Technology at Jubail Industrial College, Saudi Arabia. He has been a chartered chemical engineer for more than 30 years. He is a Fellow of the Institution of Chemical Engineers, U.K. and a senior member of the American Institute of Chemical Engineers. He holds a B.Sc. honors degree in Chemical Engineering, a Master of Science degree in Process Analysis and Development and Ph.D. in Chemical Engineering, all from Aston University, Birmingham, U.K. and a Teacher's Certificate in Education at the University of London, U.K. He has directed and conducted short courses extensively throughout the world and has been a lecturer at the university level. His articles have been published in several international journals. He is an author of six books in chemical engineering, a contributor to the Encylopedia of Chemical Processing and Design, Vol. 61, and a certified train – the mentor trainer. A Technical Report Assessor and Interviewer for chartered chemical engineers (IChemE) in the U.K. He is a member of the International Biographical Centre in Cambridge, U.K. (IBC) as Leading Engineers of the World for 2008. Also, he is a member of International Who's Who for Professionals™ and Madison Who's Who in the U.S.

Index

Also of Interest

Check out these other related titles from Scrivener Publishing

Books by the Same Author

PETROLEUM REFINING DESIGN AND APPLICATIONS HANDBOOK VOLUME 1, by A. Kayode Coker, ISBN: 9781118233696. The most comprehensive and up-to-date coverage of the advances of petroleum refining designs and applications, written by one of the world's most well-known process engineers, this is a must-have for any chemical, process, or petroleum engineer. *NOW AVAILABLE!*

PETROLEUM REFINING DESIGN AND APPLICATIONS HANDBOOK VOLUME 2: Rules of Thumb, Process Planning, Scheduling, and Flowsheet Design, Process Piping Design, Pumps, Compressors, and Process Safety Incidents, by A. Kayode Coker, ISBN: 9781119476412. The second of a three-volume set of the most comprehensive and up-to-date coverage of the advances of petroleum refining designs and applications, written by one of the world's most well-known process engineers, this is a must-have for any chemical, process, or petroleum engineer. *NOW AVAILABLE!*

CHEMICAL PROCESS ENGINEERING VOLUME 1: Design, Analysis, Simulation, Integration, and Problem Solving with Microsoft Excel-UniSim Software for Chemical Engineers Computation, Physical Property, Fluid Flow, Equipment & Instrument Sizing, Pumps & Compressors, Mass Transfer, by A. Kayode Coker and Rahmat Sotudeh-Gharebagh, ISBN 9781119510185. Written by one of the most prolific and respected chemical engineers in the world and his co-author, also a well-known and respected engineer, this two-volume set is the "new standard" in the industry, offering engineers and students alike the most up-do-date, comprehensive, and state-of-the-art coverage of processes and best practices in the field today. *NOW AVAILABLE!*

CHEMICAL PROCESS ENGINEERING VOLUME 2: Design, Analysis, Simulation, Integration, and Problem Solving with Microsoft Excel-UniSim Software for Chemical Engineers, Heat Transfer and Integration, Process Safety, Chemical Kinetics and Reactor Design, Engineering Economics, Optimization, By A. Kayode Coker and Rahmat Sotudeh-Gharebagh, ISBN 9781119853992. Written by one of the most prolific and respected chemical engineers in the world and his co-author, also a well-known and respected engineer, this two-volume set is the "new standard" in the industry, offering engineers and students alike the most up-do-date, comprehensive, and state-of-the-art coverage of processes and best practices in the field today. *COMING OUT IN SUMMER 2022!*